OCTOBER 1951

Chemical Engineering Progress

PUBLISHED MONTHLY BY THE AMERICAN INSTITUTE OF CHEMICAL ENGINEERS

TURBO-MIXER, a division of
GENERAL AMERICAN TRANSPORTATION CORPORATION

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Volume 47
Number 10

Chemical Engineering Progress

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1951

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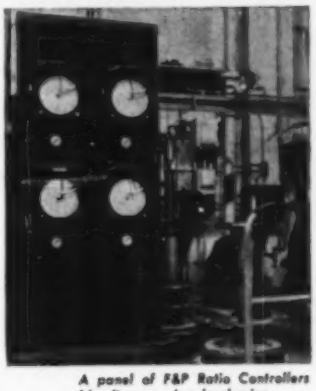

A panel of F&P Ratio Controllers blending wood pulp slurries.

AUTOMATIC FLOW RATIO CONTROLLER

A *complete* system for proportioning the flow of fluids (gases, liquids, slurries) in two or more lines. The Fischer & Porter Ratio Controller can be readily tied in with machine speeds, positive displacement meters, metering pumps or dry feeders. The elements utilize electrical, electronic or pneumatic transmission of flow in any suitable combination.

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SEND FOR CATALOG 53

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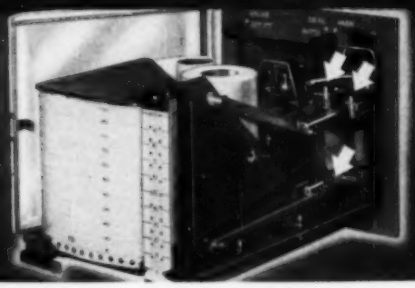
Built-in adjustable restrictions for pulsating flow



Right side of recorder, illustrating accessibility of zero and fine span adjustments.

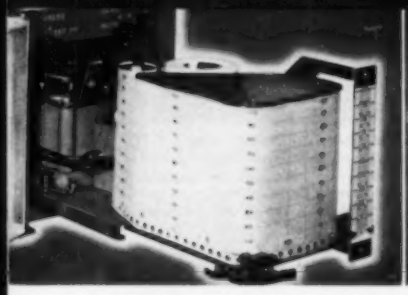
Requires very little panel space... no holes to drill

All adjustments made at front of panel

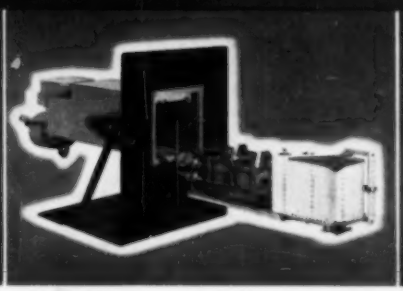
Process Control

Front view showing transparent plastic cover. Knob at bottom is for adjustment of controller set point when on "automatic," and for regulation of air-to-valve pressure when on "manual." Panel space 5 1/16" high by 5" wide.

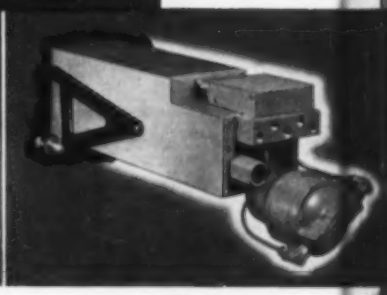
RECORDER



Left side of recorder with chassis partially withdrawn and held by safety catch. Note clearly visible 8-hour chart.



Safety catch released, chassis completely withdrawn from case. Air and electrical connections are still intact.



Rear view, showing one of the two detachable side mounting brackets with its single clamping screw.

The new *Tel-O-Set* Recorder, one of an integrated family of instruments, is used for recording such process variables as temperature, pressure, flow and liquid level. It also provides remote set-point adjustment and valve position indication.

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- ALL ADJUSTMENTS FROM FRONT OF PANEL. Chassis can be partially withdrawn. Zero, coarse and fine span adjustments, and restriction screws are accessible. Unit is in position for easy chart replacement and pen and ink service.
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- BUMPLESS SHIFTING FROM AUTOMATIC TO MANUAL OPERATION.
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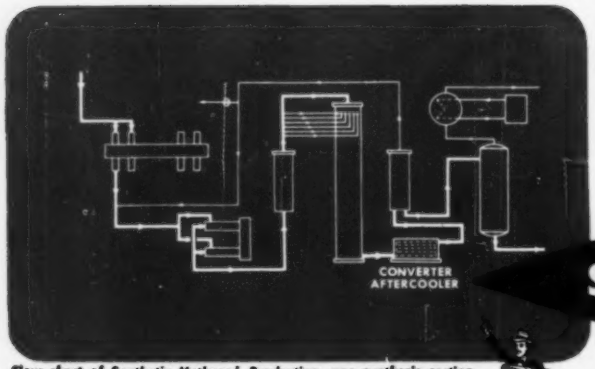
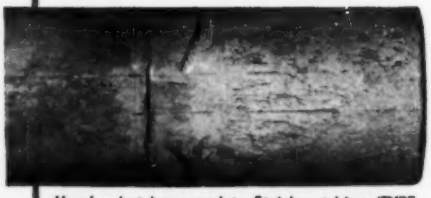
let's take a closer look at



Flow chart of Synthetic Methanol Production, gas synthesis section.

STRESS CORROSION



Here's what happened to Stainless tubing (TYPE 347) when exposed to chloride-bearing bayou water while conveying methanol and unreacted synthesis gases at 300 F and 5000 psi. Note transgranular branching cracks typical of stress-corrosion in pitted stainless steel.

Here's Croloy 5 tubing, after twelve months in the same service. Note absence of pitting and stress-corrosion. Wall thickness was not appreciably reduced.

Chloride solutions cause *pitting-corrosion* of stainless steel. Stresses on the material — whether those of construction or high-pressure operation — become localized at the root of the pits. Such stresses may reach sufficient local magnitude to cause stress-corrosion and, eventually, failure.

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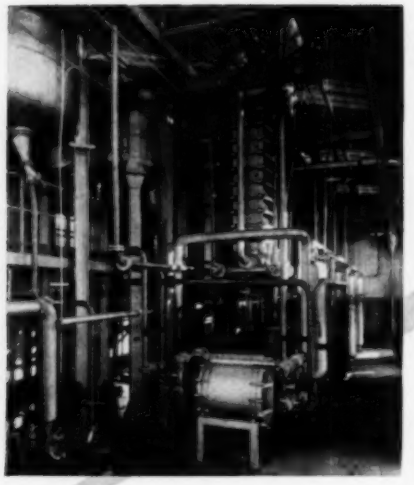

Evaporator installation with Rosenblad* channel switching system

*Patents applied for

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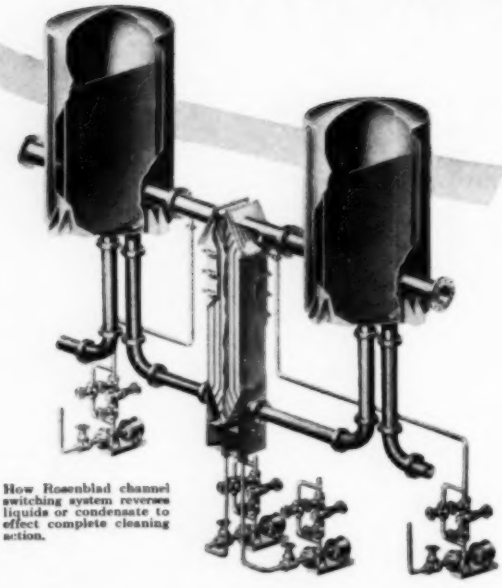

How Rosenblad channel switching system reverses liquids or condensate to effect complete cleaning action.

Now you can get *continuous full-capacity* operation . . . by using Conkey flat plate heating surface evaporators arranged in multiple effects with Rosenblad channel switching system. The Rosenblad channel switching system thoroughly cleans heating surfaces, interconnecting piping and pump systems *while in normal operation*. It eliminates expensive "down-time" or production loss for costly cleaning operations because of lime salts dropping out of solution and practically insulating the heating surface from further heat flow.

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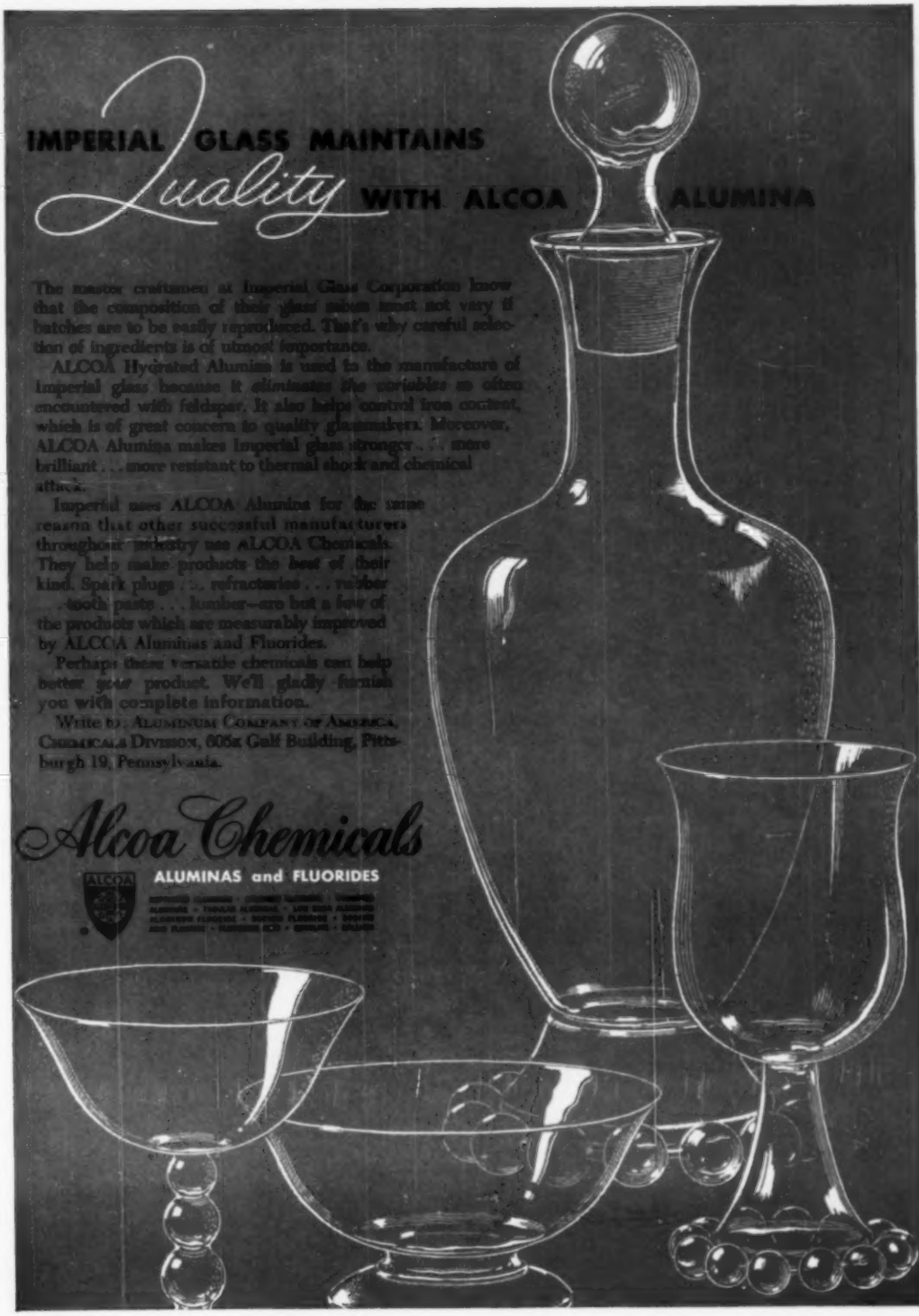
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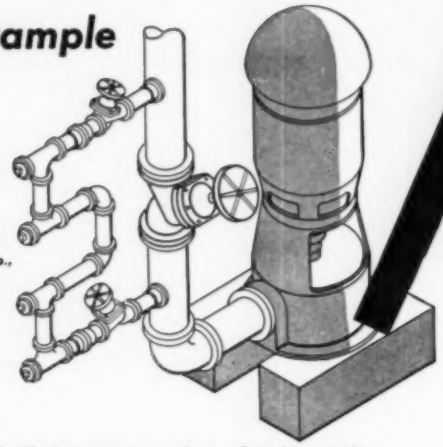

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Pounds water evaporated per pound steam input from boiler	0.9	1.75	2.5	3.2	4.0	------
Pounds steam required per 1000 gallons water evaporated	9250	4750	3325	2600	2070	200
Gallons condenser water required per 1000 gallons water evaporated	33,300	16,650	11,100	8,325	6,650	none
Steam cost per 1000 gallons water evaporated, based on steam at $.80 per 1000 pounds	$7.40	$3.80	$2.66	$2.08	$1.66	$.16
Condenser water cost per 1000 gallons water evaporated based on water at $.10 per 1000 gallons	$3.33	$1.66	$1.11	$0.83	$0.66	none
Electric Power Consumption per 1000 gallons water evaporated	10 KWH	6 KWH	6 KWH	6 KWH	6 KWH	60 KWH
Electric Power cost per 1000 gallons water evaporated based on power @ $.01 per KWH	$.10	$.06	$.06	$.06	$.06	$0.60
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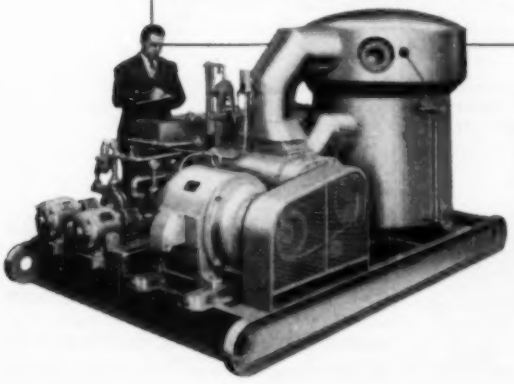

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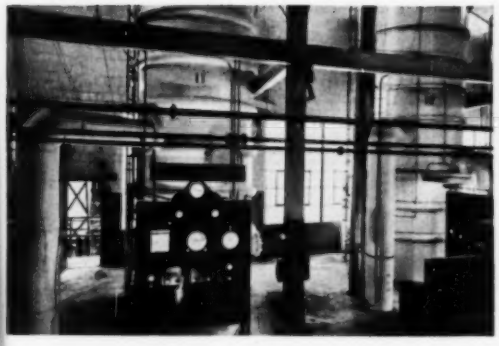

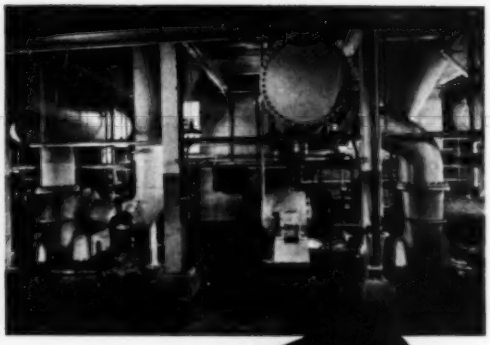

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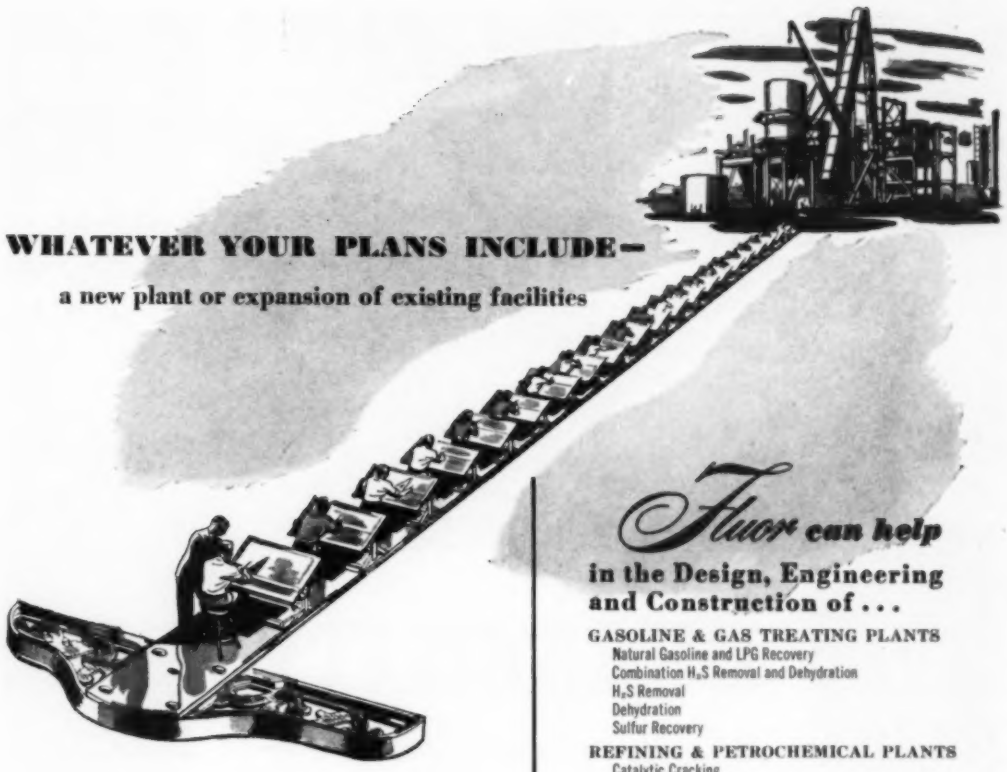

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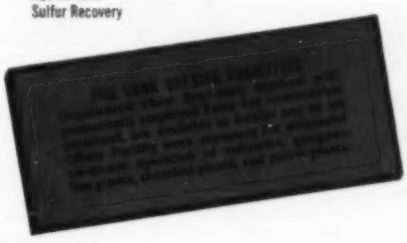

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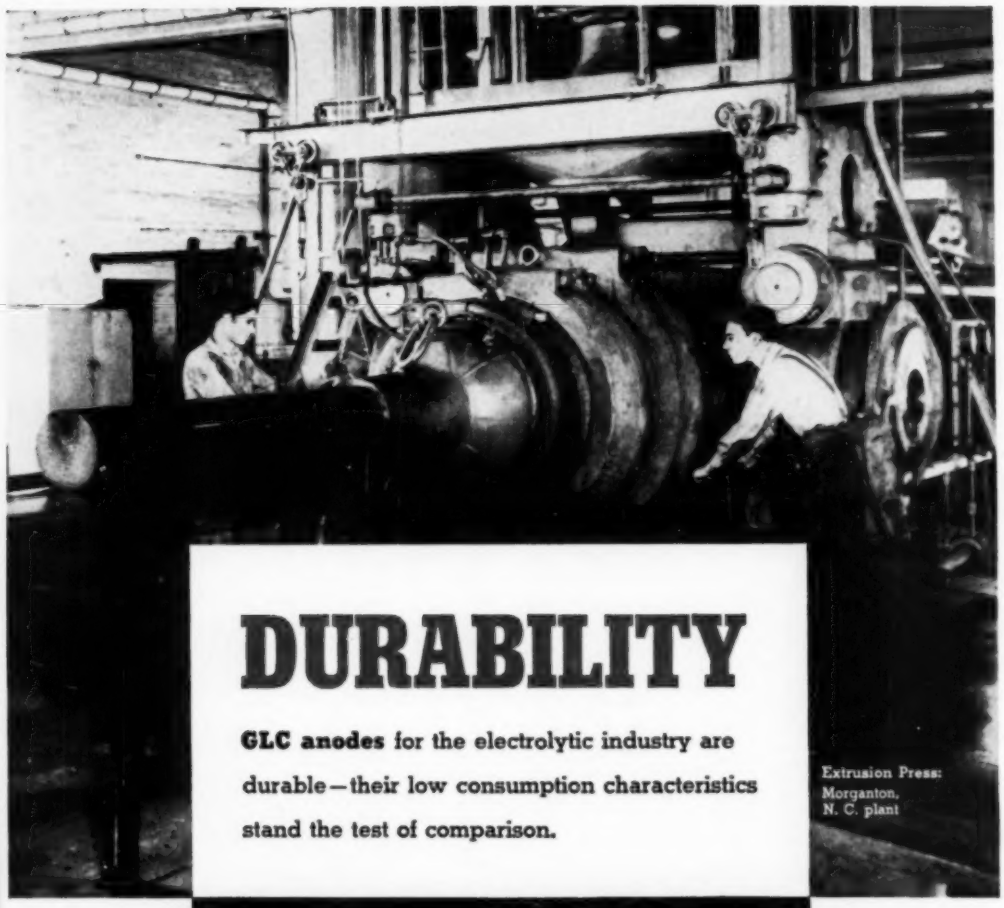

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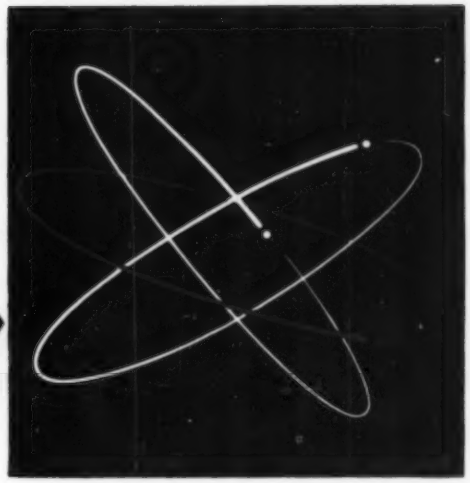

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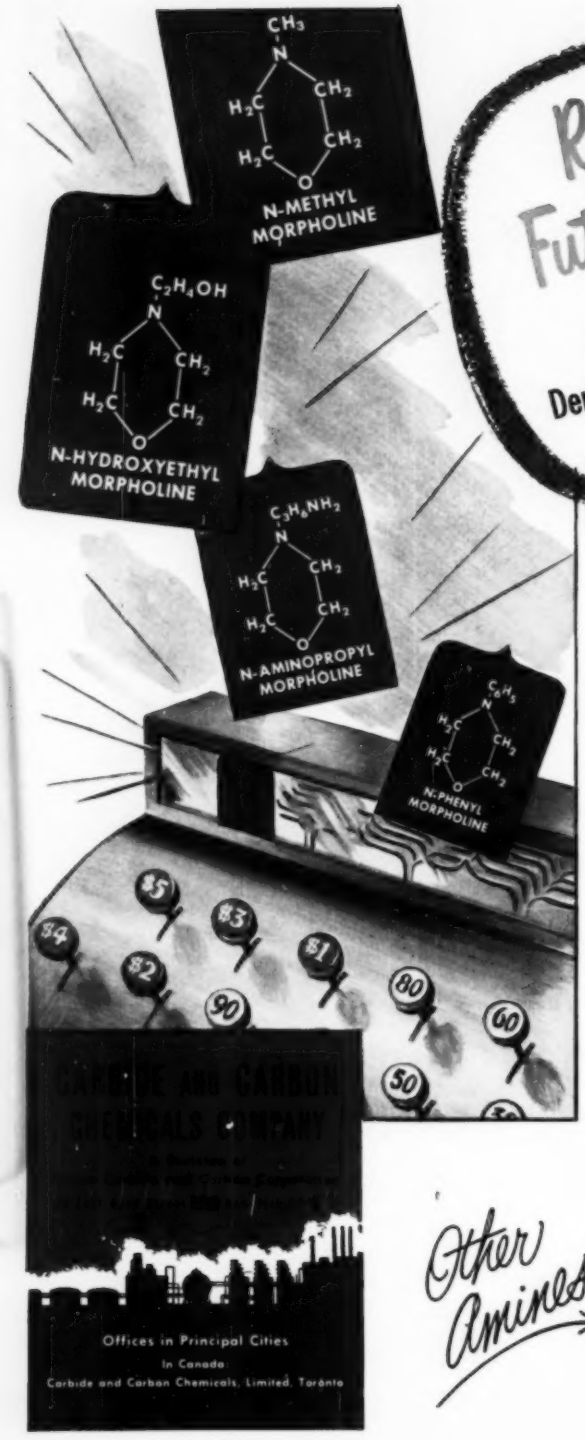

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OPINION — COMMENT

VALUES OF COOPERATION

PAUL REVERE spread an alarm, a necessary and preliminary deed to any mass action by men. Making the point of this editorial early, as necessary as Revere's ride was, the net result of the gallop would have been a sweaty horse and an aching rider if it had not been for the cooperation and likemindedness of other men. His message, terse and hurried, was understood and acted upon because the audience was prepared for action and willing to merge their personalities into one over-all and worthy cause. For success many kindred minds and bodies were mobilized.

So it is in engineering. As a calling it is individual, but it is the cooperation of many engineers in operating companies, research departments, colleges, etc., that makes for the greater success of the profession. This is an easy analogy to apply to the formation of professional societies. The mass paying of dues and the mass support by professional engineers of magazines such as this, give results which benefit all. One easy step further in this argument about the benefits of cooperation leads to the subject of its importance as a cementing medium between the professional societies themselves.

The American Institute of Chemical Engineers now has representatives on more than a dozen other professional societies working on programs of mutual concern. These are listed in the front of our directory, and it would be interesting enlightenment for all our members to note the varied fields in which the opinion and suggestions of the A.I.Ch.E. are asked. Considerably more than 40 members carry on in these positions, and through them Council is kept informed of all important problems which arise, and subsequently advise on the best course of action as it affects our profession. This is as it should be, and there is a phase of our cooperation that at this time is of immense importance, and has been of this character for several years—the supply of engineers.

Only recently has the public become aware of the facts. Fortune and the Saturday Evening Post both carried manpower stories in September. Stressed in each was the need for engineers, and yet more engineers. The public is now beginning to see the danger in a shortage of trained engineers and scientists, and a very real danger it is. But before the public can act it must be informed, and before it can be informed those who are in the best position must present the facts. In this case the A.I.Ch.E. acted, first through its representation on Selective Service Advisory Committees by S. L. Tyler, our Secretary, and later through the Engineers Joint Council and its Technical (later Engineering) Manpower Commission. This committee by gathering facts and funds from industry, was able to get across the story that the technical manpower necessary to maintain the defense effort was not available, an adequate supply was not being trained, and the men now working or about to be graduated ought to be deferred. This is an old story to our readers and we do not propose to open up the question of manpower. The point is that here is a recent example of how the engineering societies planned an action which has been successful and which gives promise of continued success in getting over to the public and to the lawmakers the facts of an engineering situation.

Such cooperation and such results are also part of the past. In 1947 E.J.C. completed a monumental study of the economic position of the engineer, and published it under the title "The Engineering Profession in Transition." In the cooperation of the five engineering societies there is a tool for good and for influence which no one of the societies alone could form or use. Perhaps here is a needed voice of authority and weight in problems affecting the engineering profession. Each society has problems peculiar to its own profession and autonomy of action for each group should be preserved. Yet it is well to know that in common problems, where concerted action is needed, the Institute is part of an organization which represents well over 130,000 professional men.

One other example of the value of cooperation lies in the Institute's work, again with other engineering groups, in the Engineers' Council for Professional Development. Formed in 1932 to enhance the professional status of the engineer, E.C.P.D., as it is abbreviated, has grown slowly. Cooperation between engineering groups was not new as an idea but it was new and strange as a program. But it kept growing. In 1932 accrediting of Engineering curricula was begun by E.C.P.D. Even then, the A.I.Ch.E. recognized concurrent effort as a force to be encouraged, and joined in this movement even though its own accrediting program, the first of its kind, had been under way since 1925.

E.C.P.D. has a fine record. In literature alone the record shows several "best sellers." Well over 180,000 copies of "Engineering as a Career" have been sold since it was first introduced in 1942; "The Second Mile" has been distributed to 77,000 engineers and students, and the "Manual for Junior Engineers," introduced a few years ago, has already had sales of 24,000.

In the future much is in store. Programs are being planned for action on the international scale with other world engineering groups, Pan American unity is being explored through the new organization U.P.A.D.I. One manpower convocation has been held by E.J.C., and others are planned. E.C.P.D. has now before it a positive program for guidance work in each local community. These plans are primarily designed to help all engineers. The A.I.Ch.E. rightly is part of them. For just as the consolidation of the hopes and plans of many individual engineers effect a successful professional society so the concerted actions of the engineering societies will make for a successful engineering profession.

F. J. Van Antwerpen, *Editor*

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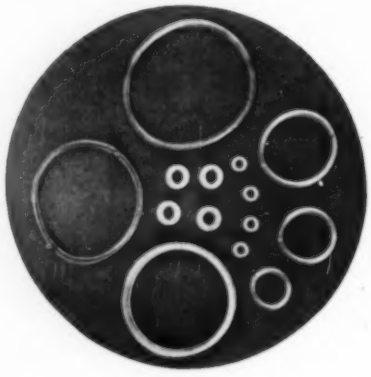

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RECENT ADVANCES IN FILTRATION THEORY

SHELBY A. MILLER

University of Kansas, Lawrence, Kansas

SHELBY A. MILLER, professor of chemical engineering at University of Kansas, Lawrence, Kan., and the author of Industrial and Engineering Chemistry's annual review of filtration, is a graduate of the University of Louisville (B.Ch.E., 1935) and the University of Minnesota (Ph.D., 1944). He has been employed by Corhart Refractories Co. (1935-36), Engineering Research Laboratory of the Du Pont Co. (1940-46), and University of Kansas (1946—date). During the past summer he worked with Monsanto Chemical Co's Engineering Sales Department.

We begin this month a series of four papers on that vital subject of chemical engineering . . . filtration . . . the first article by S. A. Miller concerns theory and the advances that have been made in engineering thinking on filtration and its action . . . the second paper authored by H. P. Grace discusses in detail the various types of filter, their advantages and disadvantages. . . .

FILTRATION is the separation of undissolved particulate or fibrous solids from a liquid by passage of the liquid through a septum impermeable to the solids. This unit operation is actually a phenomenon of unsteady-state fluid dynamics in which the resistance is variable and the flow rate need not be constant. The flow pattern, which is usually curvilinear, is sometimes partially turbulent. The pressure gradient along the fluid stream occurs by virtue of viscous friction, expansion and contraction eddies, change of the stream direction, and turbulent inertia losses. The filtration proper is often followed by washing and dewatering, operations which involve abrupt changes in the properties of the flowing fluid. As a consequence of these complex conditions, a general and rigorous description of filtration does not exist.

A reasonably satisfactory semitheoretical mathematization of filtration gradually evolved, however, during the first four decades of this century. A general rate equation was proposed

$$dV/d\theta = (\Delta P \cdot A)/(R\mu) \qquad (1)$$

where

A = area of filtration
R = resistance identified with filtration
V = total volume of filtrate collected
ΔP = pressure drop across filter
θ = duration of filtration cycle
μ = filtrate viscosity

It was recognized that in the usual filtration, the major resistance lay in the filter cake and therefore increased as the cake grew. Sperry (*25*) early proposed that the resistance was in two series parts, that of the filter cake (R_K) and that of the filter medium (R_F), so that

$$dV/d\theta = (\Delta P \cdot A)/\mu(R_K + R_F) \qquad (2)$$

The cake resistance R_K could be expected to vary with the thickness of the cake and hence with the quantity per unit area:

$$R_K \propto W/A = \alpha W/A$$

where

W = mass of cake formed

The proportionality constant α, called the specific cake resistance, often depended upon the pressure to which the cake was subjected (several relationships defining α as a function of filtration pressure were proposed); furthermore, it differed not only for each material but also even for batches of the same material which had undergone different treatment.

The filter medium resistance R_F was found also to depend on pressure, but not usually in the same way as did α. It was soon discovered that the effective filter medium resistance was far different from that of the bare, clean cloth and that the value of R_F needed to be determined, therefore, in the course of an actual filtration.

Meanwhile, a second broad type of filtration was defined (*11*), one in which no apparent cake formed, but solids were collected within the body of a filter medium. Whereas for cake filtration, the total resistance was linear with respect to the volume of filtrate processed, in filter medium filtration the resistance increased more than linearly with V, the specific relationship depending on the mechanism of cloth blinding, as will be discussed in a later paragraph. The magnitude of the resistance, however, depended also on the particular material being filtered.

Detailed, authoritative reviews of the development of the filtration rate equations and of the concept of filtration resistance have been published by Carman (*7*) and by Heertjes (*9*).

Although the rate equations which evolved were extremely useful in shedding light on the mechanism of filtration and in guiding the scale-up of laboratory data to plant-size filters, they were unable to define in fundamental terms the resistance coefficients which they contained. Thus the resistance was almost impossible to predict, its nature resisted delineation, and it was even difficult to reproduce.

Since 1946, a number of mathematical and experimental investigations have been directed specifically toward a more precise definition of filtration resistance. There have been three principal approaches to the problem: (1) determination of the mean or local specific resistance of a filter cake in terms of its permeability and the properties which affect permeability; (2) examination of the resistance of the solid layer adjacent to or perhaps incorporated in the filter medium; (3) description of the resistance of a filter cake, by the principle of dimensional similitude, as the friction loss experienced by the fluid which passes through the cake. The work in each of these directions will be discussed briefly.

Nature of Specific Resistance

Both Ruth (*21*) and Carman (*7*) showed that the mean specific resistance of a given cake is a function only of the pressure applied to the cake. Although the point resistance α_p varies through the thickness of a cake of compressible solids, the mean, defined

$$\alpha = P_f / \int_0^{P_f} (dP/\alpha_p) \qquad (3)$$

depends only on the filtration pressure, if α_p is a function only of P. This is a conclusion of tremendous consequence, for it opens the way to a definition of the resistance encountered in filtration in terms of the filtering pressure

(whether constant or variable) and the properties of the cake aggregate.

Meanwhile investigators of fluid flow through granular beds had advanced to a similar point of defining bed permeability in terms of particle and bed properties. It is not surprising, therefore, that those interested in filtration turned their attention to the existing correlations of granular bed permeability, particularly to that of Kozeny (*16*). Kozeny's equation may be written

$$u = \frac{dV}{Ad\theta} = \frac{g_c\epsilon^3}{5(1-\epsilon)^2S^2\mu} \cdot \frac{\Delta P_c}{L} = K_k \frac{\Delta P_c}{\mu L} \qquad (4)$$

where

g_c = gravitational conversion factor
K_k = Kozeny bed permeability
L = bed thickness in direction of net fluid flow
P_c = fluid pressure within bed
S = specific surface, area per unit volume of solids
u = superficial velocity through bed, based on bed cross section
$\Delta P_c/L$ = pressure gradient in fluid through bed
ϵ = bed porosity, volume of voids per unit volume of bed

Carman (*7*) pointed out the similarity between Kozeny's equation and the differential rate equation for filtration

$$dV/d\theta = (\Delta P_c \cdot A)/(\mu R) \qquad (5)$$

If the total resistance R is defined in terms of a mean specific cake resistance α per unit mass of cake of unit cross section,

$$R = (\alpha Vc)/A \qquad (6)$$

and since

$$Vc/[A\rho_s(1-\epsilon)] = L \qquad (7)$$

$$R = \alpha L\rho_s(1-\epsilon) \qquad (8)$$

The filtration equation then becomes

$$dV/d\theta = (\Delta P_c \cdot A)/[\mu\alpha L\rho_s(1-\epsilon)] \qquad (9)$$

whence, by comparison with Kozeny's equation

$$\frac{1}{K_k} = \alpha_k = \alpha\rho_s(1-\epsilon) \qquad (10)$$

Thus, the specific filtration resistance should be calculable in terms of the porosity and interstitial surface of the filter cake

$$\alpha = \alpha_k/[\rho_s(1-\epsilon)] = [5(1-\epsilon)S^2]/\rho_s\epsilon^3 g_c \qquad (11)$$

One should remember that a particular compressive stress for which the resistances are valid is implicit in Equation (11). For the values of α and α_k to be comparable generally, the filtration pressure and the maximum compressive stress which the permeability bed has experienced (either in the permeability test proper or during its previous history†) must be the same, unless the cake is practically incompressible.

One should note also that the porosity value (and possibly the specific surface) to be employed is also a mean and must be determined over the entire thickness in the direction of the fluid flow. The local or point value of ϵ will vary to correspond with the magnitudes of α_p and S. If the effective specific surface does not vary with increasing compression, Equation (11) predicts a constant value of the ratio $(1-\epsilon)/\epsilon^3\alpha$ for a given cake regardless of the pressure, as suggested by Walas (*27*). Walas filtered nine materials of variable compressibility and specific resistance at pressures of 10 to 50 lb./sq.in. For a given material the values of $(1-\epsilon)/\epsilon^3\alpha$ deviated as much as 38% from the mean for that material, admittedly not a great amount in contrast with the 130-fold range of magnitude for all the materials. Walas proposed that the greatest deviations may be the result of sedimentation in the filter; but reduction of effective surface as the cake is compacted, reflected in an apparent decrease in the ratio with increasing pressure, also may have been involved.

Hoffing and Lockhart (*13*) tested the correspondence between α and α_k by conducting filtration and water permeability experiments with virtually incompressible cakes of sized quartz, diatomaceous earth, and mixtures of the two. They limited their experiments to a single low-pressure differential of 10 lb./sq.in. The observed deviation between the values of α and α_k, less than 5%, was well within the limits of the reproducibility of the measurements.

Ruth (*23*) studied the permeability of beds of carefully packed uniform spheres and derived the semitheoretical equation

$$\frac{dV}{Ad\theta} = \frac{\nu^2\rho_s f^2 D_p^2}{K} \cdot \frac{A\Delta P_c g_c}{\mu W} \qquad (12)$$

where

D_p = particle diameter
f = shape factor, ratio of specific surface of a sphere to specific surface of particle in question
K = constant
W = mass of solids comprising bed
ν = specific void volume, void volume per unit volume of solid

† If the cake is elastic, it will not retain the resistance imparted by previously applied stress.

The shape factor is the one proposed by Kozeny (*16*) and Equation (12) is, in fact, an empirical modification of Kozeny's equation. It may be written in terms of S instead of fD_p. Application of Equation (12) to cakes of sub-sieve-size particles was not satisfactory, however: not only did the shape factor computed by the equation from permeability data decrease with decreasing size of particles of the same material, but also the factor was always smaller than one would predict from microscopic examination.

Ruth's explanation of the failure of Equation (12) was that a portion of the void space is unavailable for fluid passage, and that the equation should be written

$$\frac{dV}{Ad\theta} = \frac{(\nu-\nu_0)^2\rho_s f^2 D_p^2}{K} \cdot \frac{A\Delta P_c g_c}{\mu W} \qquad (13)$$

In a clever experiment, in which he varied porosity by mechanical stress artificially imposed upon the cake and observed the consequent permeability, he supported his explanation and inferred the magnitude of ν_0 (Fig. 1).

Although unavailable void volume may reside in blocked-off pockets or completely occluded chambers and may result from molecular adsorption of fluid on the interstitial walls, Ruth's hypothesis of an electrically adsorbed ion layer and the consequent phenomena of streaming potential and electroosmosis appears to account most plausibly for the decreased effective void. The exposition of his electrokinetic thesis is recommended reading (*23*).

Equation (13) may be used to define the specific cake resistance

$$\alpha_p = (\Delta P_c A^2 d\theta)/(\mu W dV) = K/[(\nu-\nu_0)^2\rho_s f^2 D_p^2 g_c] \qquad (14)$$

where α_p is the point resistance characteristic of the particular mechanical stress imposed during the permeability test. If α_p is determined at several pressures, the mean specific resistance may be computed from Equation (3). Ruth so computed the cake resistance and compared it with values calculated from filtration tests which, unfortunately, had been made at a considerably earlier date. The elapse of time between filtration and permeability determinations casts doubt on a comparison of the two; even so, the two sets of resistance data differed only by 10 to 15%.

Walas (*27*) also formulated an expression for α from a statement of Kozeny's equation. On the assumption of a substantially invariant shape fac-

tor, he concluded that the proportionality

$$(1-\epsilon)/(\epsilon^3\rho_s a) \propto D_p^2 \qquad (15)$$

should obtain for all the materials which he investigated. Experimentally, however, he discovered that the exponent of D_p was 1.02 instead of 2. A possible explanation for this result, as he suggested, lies in an unavailable void volume; the assumption of constancy of shape factor may also contribute to it. In view of the confirmation of the second power by other investigators (*13, 23*), Walas' correlation is not recommended.

Equations (11) and (14) define the specific filtration resistance in terms only of particle and cake properties. If one can determine the porosity or void volume as a function of pressure, the particle diameter and shape factor or specific surface, and the particle density, he may compute the filtration resistance. The void volume may be determined by the compression cell described by Ruth or, for limited pressures, by the Hoffing-Lockhart permeability cell. Specific surface may be determined by a permeability test. Particle diameter is less easily determined. Sedimentation and screening are unreliable for irregularly shaped particles, and microscopic examination, tedious at best, becomes unsatisfactory for such irregular particles. All three methods fail for a particle of appreciable internal surface —an effective diameter is of little practical significance, for example, with reference to a diatom (*13*). Finally, the problem of a mean particle size to be assigned to a mixture containing more than one size, the problem which for so long clouded the analysis of flow through granular beds, still has no acceptable solution. The harmonic mean suggested by Ruth and by Walas is unsubstantiated for wide extremes.

For these reasons, the specific surface appears to be the wiser correlating variable. Determined, along with available void volume, in a simple compression permeability cell, it offers an attractive design and control criterion.

The Initial Layer of Cake

The recent study of filtration resistance as related to cake permeability is not helpful in an evaluation of that portion of the resistance which resides outside the cake proper, principally in the filter medium and its support. A complete description of the resistance in a filter will require a specification of external resistance in addition to the specific cake resistance, whatever the method of estimation of the latter. Theory has not yet provided a means of predicting the effective filter medium resistance; it must be determined *in situ* during an actual filtration test.

Nevertheless, considerable attention has been directed to the filter medium and its permeability. In 1935 Ruth (*22*) summarized the investigations of filter medium resistance to that date and demonstrated by use of his own data and those of others that the flow through the medium at all times during a normal filtration cycle is streamline. Without such assurance, one could not generally expect filtration to be a Poiseuille phenomenon from its very outset. The permeability or resistance of a clean, new cloth is not useful for the estimation of cloth resistance. Not only does the cloth resistance increase with continued use (many industrial fabrics fail finally because of blinding), but it may become temporarily much greater while it is supporting the filter cake (*22*).

The latter observation directs attention to the boundary layer of solids, and led Heertjes and Haas (*10*) to undertake an analysis of the initial stages of constant-pressure filtration.

These investigators, conducting a study that was principally experimental, worked with extremely dilute slurries (maximum solids concentration of 4g./l.) in an attempt to stop a filtration after only a very thin cake had been formed and yet have sufficient time to observe accurately the character of the time-discharge data under a given pressure difference. Their approach can be criticized on the grounds of relevance to the problem originally proposed, but the results of their careful study shed valuable light on true filter-medium filtration and provide interesting inferences about the mechanics of the deposit of a cake on a septum.

Some years ago Hermans and Bredée (*11*) proposed that filter-medium filtration should occur by one of two distinct processes: (1) individual solid particles completely block individual pores in the medium (the case of complete blocking, wherein the change of resistance with respect to filtrate volume increases as its own square); (2) particles gradually collect on the walls of the capillaries in the medium and ultimately fill the capillary (the case of semi-blocking, wherein the change of resistance with filtrate volume increases as the three-halves power of resistance). They found an additional empirical case in which the increase of resistance with filtrate volume was proportional to itself; for this case, they could propose no physical explanation.

Heertjes and Haas pursued the idea of Hermans and Bredée, and derived a rate equation for complete blocking:

$$dV/(Ad\theta) = \Delta P_F \Big/ \left[\frac{R_F W}{aN - W} + R_F\right] \qquad (16)$$

where

a = mass of solids which will block a single pore
N = number of pores per unit area of filter medium
ΔP_F = pressure drop across the filter medium
$R_F W/(aN-W) = R_K$ = resistance of the "cake"

They pointed out that the cake resistance is conceptual rather than actual, since it represents simply increased diaphragm resistance.

The Heertjes-Haas data proved to be most interesting. They showed that with decreasing prefilt concentration, the character of cake resistance shifted from that of cake filtration ($R \propto V$, or a constant) through the gamut of Hermans-Bredée laws to approach the complete blocking case (Fig. 2). Heertjes (*9*) concluded that the various forms of filter-medium filtration do not represent a family of discrete steps, but rather a continuous spectrum of relationships bounded by the limits of cake filtration and complete blocking. It is probable that position within this spectrum is determined not only by the slurry concentration, but also by the par-

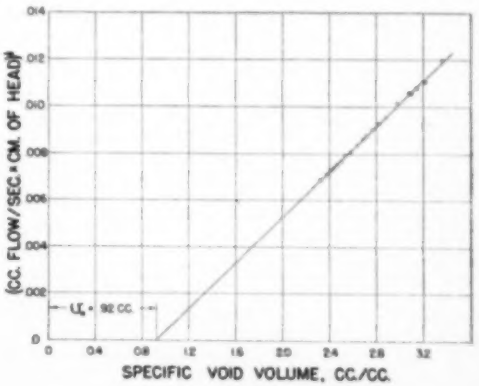

Fig. 1. Determination of dead void volume in a permeability-compression test. (Data and method of Ruth)

ticle character and the nature of the filter medium as well.

A paper by Ishikawa (*15*), only recently abstracted in this country, appears to analyze by theory the boundary layer of solids on a filter medium and to report permeability relationships that depend on the character of the solids and of the medium. These data may well support the hypothesis of Heertjes and Haas, but a definite conclusion must await the arrival of the original paper.

Friction Drop through Filter Cake

The flow of filtrate through a cake is a phenomenon of fluid dynamics wherein the pressure energy of the upstream fluid is dissipated against frictional, eddy, and kinetic energy losses encountered in its passage through the granular aggregate which is the cake. In general the flow is laminar and the energy dissipation is one of viscous friction. One would expect, therefore, that relations between flow rates, pressure gradient, and fluid properties could be correlated for a given bed (as for a pipe of given cross section and roughness) by a modified friction-factor chart. Such, of course, is the case, and for a number of years friction-factor correlations for pressure drop across beds of packing have been employed with moderate success. Almost 15 years ago Carman (*6*) proposed a modified plot of friction factor vs. Reynolds number based on the specific surface of the bed material and modified by a porosity term.

The newest attack of filtration resistance, involving a semitheoretical correlation similar to Carman's, is due to Brownell and Katz (*1, 2*), who developed a modified Moody chart for beds of particulate materials. The chart coordinates are a Reynolds number based on a particle diameter (mean surface diameter) and the superficial fluid velocity through the bed, and modified by the factor $1/\epsilon^m$; and a friction factor based on the same diameter and velocity and on the bed thickness, and modified by the factor ϵ^n. The exponents m and n are complex functions of the bed porosity and the particle sphericity.

Because the modified coordinates become the conventional moduli when porosity becomes unity (i.e., in an empty pipe), Brownell and Katz claim that m and n can be so selected as to cause the data for granular beds to coincide with those for circular pipes. This they have done with some degree of success, although the data deviate in the critical range and the roughness coefficients assigned to the various particles tested do not provide a sharply defined set of parameters in the turbulent region.

In a second paper, Brownell and Katz made a more significant contribution

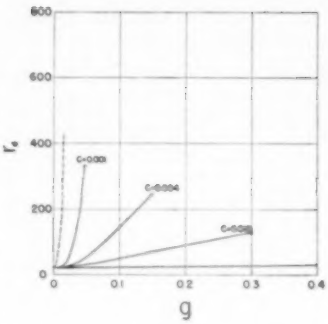

Fig. 2. Differential specific resistance (r_s) of purified as a function of amount of substance (g) on nylon cloth.

(Data and correlation of Heertjes and Haas)

(*1, 3*). In an analysis of the displacement of one fluid from a granular bed by a second fluid immiscible with the first (as in the air-blowing of a filter cake), they developed a method of estimating the unsteady-state rate of discharge of both fluids and the pressure drop identified with their discharge. The computations required are lengthy; the correlations upon which they are based are spotty; and the quantities involved in the calculations are difficult to determine and critical enough to make their rough estimation undesirable (e.g., particle diameter, bed porosity, particle sphericity, residual saturation of the bed). In spite of these disadvantages, the method proposed permits the first attempt to estimate such important items as amount of air required for the blowing of a filter cake and the moisture content of the cake after blowing. It constitutes a significant contribution to filtration and it is worth the attention of other investigators interested in the field.

Lapple (*17*) criticized the friction-factor plot of Brownell and Katz for its lack of proved generality and its consequent inaccuracy for high-porosity aggregates, particularly in the turbulent region. Carman's correlation, in his opinion, is more sound theoretically and more reliable. Although Lapple's comments appear justified, and one should use the friction-factor plot with caution for conditions not covered by the data correlated originally, the plot probably can be used with safety in the viscous section unless particularly unisometric particles are involved. A more serious practical deterrent lies in the difficulty of particle size and shape determinations, porosity and residual saturation estimations, and labor of calculation. These disadvantages virtually prohibit use of the friction-factor method for estimation of conventional filtration resistance for all except the rare cases of transition or turbulent flow. With the data needed to calculate filtration resistance from the Brownell-Katz chart, one can make the computation more easily and with greater assurance by one or another of the existing permeability relationships. Two-phase flow is another matter, for in this area the Brownell-Katz correlation provides a badly needed tool.

Brownell and his co-workers have supplied several examples of the use of their method for predicting filter design, dewatering-air requirements, and the pressure drop across a cake on a filter-dryer (*4*). Their correlations with examples of their uses also are presented in Brown's textbook on unit operations (*1*).

Experimental Validation and Application of Theory

Slowly and painfully through the years in which the filtration equations were formulated, data were accumulated to verify the proposed relationships, to define their limitations, and to specify the numerical value of their constants. The validity of applying results obtained with laboratory equipment to the plant was established. Continuous vacuum filters were shown to be no different in principle from batch pressure filters, and data obtained with one type were proved to be applicable, within bounds, to the other. The filtration rate equations were employed, although not to their limit, to predict the effect of certain process changes on the filtration cycle and to define desirable operating conditions (e.g., the optimum relative filtering and washing cycles). All these contributions were made in a period ending well over ten years ago and need not be specifically cited here.

In the past decade, the value of using small, precise equipment to obtain fundamental filtration data and to supply necessary design information has been emphasized. Sperry (*26*) and Hoffing and Lockhart (*13*) described laboratory pressure filters that require relatively small amounts of slurry and that are capable of much greater precision than the usual semiworks filter. Both reported typical data obtained with their apparatus.

The most significant recent validating data have been those of Ruth (*23*) and of Hoffing and Lockhart (*13*), already discussed, which support the application of permeability observations to filtration. Of particular interest is Ruth's report (*23*) that mechanical stress imposed upon a preformed filter cake will indicate accurately the compressive effect of friction drag experienced by the cake during its formation.

One of the most valuable applications of the filtration equations is their use to predict quantitatively the effect of a change of process conditions on a filtration cycle. Mondria (*19*) and Reeves (*20*) have so applied the constant-pressure equation to the analysis of continuous vacuum filters. Mondria wrote the equation in a form that permits easier prediction of the effects of drum speed and of cloth blinding on filtration rate. The effects are relative to some original condition and rate and can be determined without the necessity of evaluat-

ing cake resistance and filter-medium resistance. Reeves simplified the rate equation by assuming negligible filter-medium resistance and used it to formulate expressions for optimum drum speed, optimum dilution of a prefilt of relatively high filtrate viscosity, and maximum amount of wash liquor that can be applied to the cake under various conditions of solvent dilution.

Hill and Kaiser (*12*) stated that the effect on vacuum filtration rate of a change in slurry temperature can be estimated from the filtration equation, since the major influence of temperature change will be on filtrate viscosity. Data which they report, however, show a greater effect than can be accounted for by viscosity change alone, and they illustrate the danger of oversimplifying the importance of an implicit process variable.

The evaluation of constant-rate filtration data is unpleasant because of the simultaneous calculation of the four necessary constants. A recent paper (*14*) presents a simplified approximation of the integrated constant-rate equation:

$$\Delta P_o = a_o(\Delta P_o)^b(\mu c/\rho)(W/A\theta)^2\theta \qquad (17)$$

where

- b = compressibility coefficient, constant for a given slurry
- c = slurry concentration
- a_o = cake resistance coefficient, constant for a given slurry

A major contribution of the paper is the suggestion that the pressure at the base of the cake is greater than the pressure drop across the septum at the start of filtration and may be estimated by trial and error. Although this equation is much simpler than any previously proposed, it needs further validation by experiment to warrant its general use.

Burak and Storrow (*5*), following an earlier suggestion of Maloney (*18*), have embarked on an intriguing study of the analogy between centrifugal and conventional filtration. Space will not permit a review of their work at this time and any serious criticism of the work should await the publication of promised subsequent papers on this subject by Professor Storrow. Suffice it to report here that the permeability of cakes of the same material formed respectively in a filter and in a centrifuge differ widely and that the fact of centrifugal stress does not seem to account entirely for the difference. The equations developed by Burak and Storrow are worth considerable thought; it appears that they may not be completely general. Storrow is to be commended for undertaking this complex study and for the care with which it has been started. Further results from it are anticipated with interest, for the investigation promises to be rewarding.

Theory of Filtration

A pure theoretician may offer the criticism that filtration theory is really not theory at all, but a piece of reasonable mathematization, at best semi-theoretical, which helps explain the operation. It may be well, therefore, to conclude this paper with a reference to a soundly rigorous approach to filtration. Ferrandon (*8*) made a classical analysis of the hydrodynamics of flow of a perfect fluid through a porous mass. The result was a set of tensors said to characterize the permeability and wetted surface of the porous mass—and they are completely unsolvable in terms of numerical values for the properties of a given actual system! One may agree with the theoretical critic and experience discouragement at the results of Ferrandon's effort. Or on the other hand, he may concur with Ruth, who declared 20 years ago in the first (*24*) of his series of notable papers on filtration: "It can be said with assurance that no other branch of chemical engineering lends itself more exactly to theoretical treatment and mathematical analysis and computation than does this much maligned unit process of filtration." In either case, he must acquiesce that the mathematical relationships which today are called the theory of filtration aid materially in the comprehension of this important operation and constitute a sound, reliable guide to its effective practice.

Notation

- A = sectional area of a filter, filter cake, or granular bed transverse to net direction of fluid flow
- a = mass of prefilt solids required to block a single pore of filter medium
- b = compressibility coefficient of a cake
- c = slurry concentration
- D_p = diameter or characteristic dimension of a particle
- f = shape factor, ratio of specific surface of a sphere to specific surface of a particle
- g_c = gravitational conversion factor
- K = constant
- K_k = Kozeny permeability coefficient $= \epsilon^3 g_c/5(1-\epsilon)^2 S^2 \mu$
- L = thickness of a filter cake or granular bed in direction of net fluid flow
- m, n = dimensionless numbers, functions of porosity of a granular bed and sphericity of particles comprising it
- N = number of pores per unit area of filter medium surface
- P_c = fluid pressure at any point within a filter cake or granular bed
- P_f = filtration pressure $= P_c - (P_c)_2$, where $(P_c)_2$ is the fluid pressure at base of filter cake
- ΔP = pressure drop across a filter
- ΔP_F = pressure drop across a filter medium
- R = total filtration resistance $= \Delta P/(\mu dV/Ad\theta)$
- R_F = resistance of filter medium
- R_K = total resistance of a filter cake or granular bed
- S = specific surface of a particle or mass of particles, surface area per unit of particle volume
- u = superficial velocity through a filter cake or granular bed based on the total cake or bed cross section $= dV/Ad\theta$
- V = total volume of filtrate collected in time θ
- W = total mass of solids in a filter cake or granular bed
- a = mean specific resistance of a filter cake, unit mass basis, $= RA/W$
- a_o = cake resistance coefficient
- a_k = specific resistance of a granular bed $= 1/K_k$
- a_p = point or local specific resistance of a filter cake
- ϵ = porosity of a filter cake or granular bed, = void volume per unit volume of cake or bed
- θ = duration of a filtration cycle
- μ = absolute viscosity of a fluid
- ν = specific void volume, void volume per unit volume of solids
- ν_a = unavailable specific void volume
- ρ_s = density of a particle or of the solids comprising a cake or bed

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WHAT TYPE FILTER AND WHY

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H. P. GRACE has spent the past five years at the Engineering Research Laboratory of the Du Pont Co. This period has been occupied by research and plant problem work on mechanical separations, particularly filtration and centrifugation, and forms the background for much of the material contained in the accompanying article. Mr. Grace went to Du Pont after he was awarded a B.S. in chemical engineering by the University of Pennsylvania in 1941, and was employed by the engineering department. Three years were spent as a field engineer at the company's Baltimore (Md.) plant where much practical knowledge was gained in the filtration of pigments. Then came two years of naval electronics experience.

TO answer the question in the title requires for this paper the assumption that chemical engineers are generally familiar with the various types of filters. This paper will attempt to point out the fields of application for various basic filters, as well as their relative advantages and disadvantages, without describing in detail either their mechanical design or operation, except when unusual features exist.

Miller (*1*) discusses the basic difference in both mechanism and theory between the cake-filtration and solution-clarification types of filtration. This basic difference is important in the practical application of commercial filters since most units are best suited to either cake filtration or solution clarification while only a few must be considered for both types of operation. In Figure 1, separation by filtration has been divided into these two basic types of operation with the general-selection criteria and general-filter types outlined for each type of service.

For this discussion, an arbitrary concentration range, typical of each type filtration, will be assumed. Thus, in solution clarification the solids content is nearly always less than 0.1% by volume and may be as low as a few parts per million. In cake filtration the solids content is nearly always above 1% solids by volume and is often in the range of 3-20% by volume.

The assumed concentration ranges typical of each type operation do not overlap, and a concentration range of 0.1 to 1.0% solids by volume is not included in either. This is a region where the mechanisms of both types of operation are significant and is a region to be avoided. Although the equipment and methods suitable for either cake filtration or solution clarification can be applied to suspensions in this concentration range, neither is generally satisfactory nor economical. Thus, when faced with feed suspensions with 0.1% to 1.0% solid content, concentration by thickeners before filtration may be the best answer.

Fig. 1. Separation by filtration.

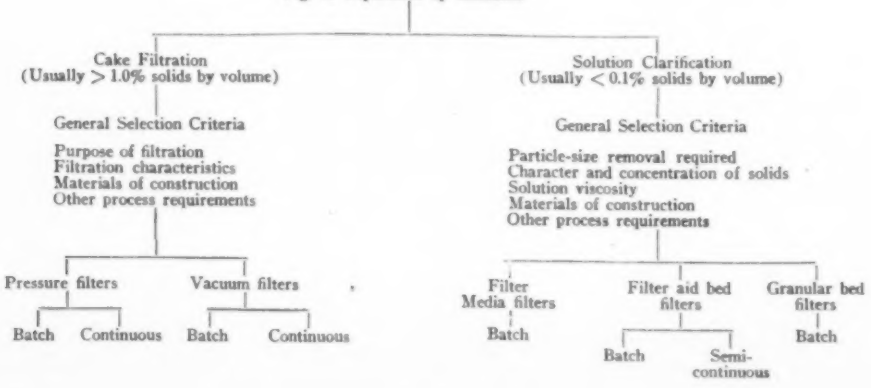
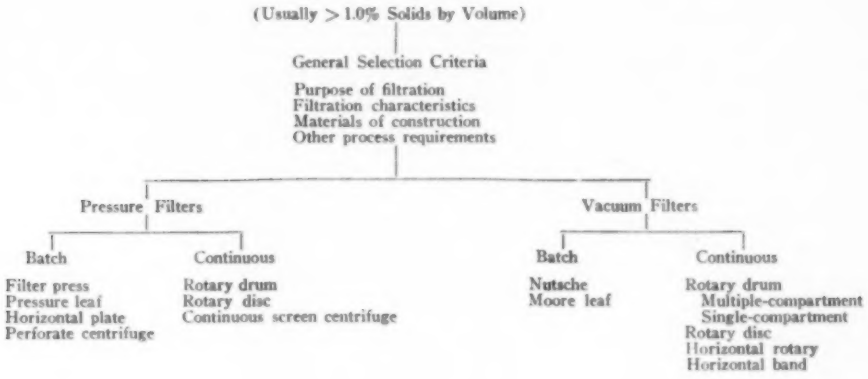


Fig. 2. Cake filtration.



General Selection Criteria

Criteria of major importance in determining the best filter for either cake filtration or solution clarification have been generalized in Figure 1 and must be considered as the next step in the selection procedure. In cake filtration, the purpose of the filtration may be (1) recovery of valuable cake solids, either highly washed or highly dewatered, (2) recovery of strong filtrate with cake solids discarded, or (3) recovery of both pure solids and strong filtrate. Thus the purpose narrows the field of selection. A filter giving good over-all washing efficiency and dry-cake discharge is important for recovery of pure solids; on the other hand a filter giving an efficient displacement wash with minimum filtrate holdup as well as reslurried cake solids is best adapted for strong filtrate recovery.

In cake filtration the filter cycle is controlled largely by the resistance characteristics of the cake solids, and the selection between pressure or vacuum operation depends largely on quantitative determination of the filtration characteristics by laboratory methods. A knowledge of filtration characteristics, including specific cake resistance and cake porosity (determined from laboratory filtration rate tests) permits calculation of results of typical operation on various commercial filters, and is essential to intelligent filter selection for cake-filtration service. For instance, materials with a specific cake resistance greater than 5×10^{11} ft./lb. are usually best handled by pressure filtration unless they are highly compressible.

Materials of construction may be a limiting factor in either type of filtration. This must be considered as it influences both filter design and cost as well as the filter medium material which can be used. Although the synthetic fabrics available today provide a wide range of chemical and temperature resistance, some conditions still exist for which suitably resistant filter media are not yet available in a convenient form.

Other process requirements such as capacity, operating temperatures, whether process is continuous or batch, whether volatile or toxic liquid is involved, are important in selecting either cake-filtration or solution-clarification equipment. In cake filtration the capacity requirement, in combination with the method of processing, usually determines whether a continuous filter will be practical and economical. The handling of volatile or toxic liquors may dictate use of a totally enclosed filter and thus eliminate consideration of other types. Operating temperatures may indicate a jacketed filter, again limiting the selection, and also operating temperatures may severely limit the type of filter media usable. Temperature may also be a serious limitation to continuous vacuum filtration since the usable vacuum may be limited by flashing.

In solution clarification the purpose is limited to recovery of clarified liquor in nearly all cases and is therefore not a variable in filter selection. Because particle removal occurs largely by entrapment within the filter medium or bed, however, the particle size to be removed and degree of particle removal are important in selection, since performance depends largely on the filter medium or bed used and not on formation of a filter cake. Thus, various filter types for solution clarification can best be classified by the type of filter medium or bed they employ. Size and degree of particle removal are real considerations in selection of the actual filter type. In the same manner, the character and concentration of the solids present are vital to selection since certain filter media or beds are best adapted and most economical only over limited ranges of solids concentration. Many media or beds are incapable of efficiently removing solids below a given particle size and knowledge of this is necessary for filter selection. Although solution viscosity is an important operating variable in cake filtration, it is usually not essential to filter selection in this type operation as it seldom varies widely from water viscosity. In solution clarification, however, solution viscosity may be often as high as 50 to 100 poises and in some cases may be greater than 1000 poises. Thus, solution viscosity is a consideration in solution clarification as it severely limits the filter design which should be selected.

Cake Filtration

In Figure 2 are shown general selection criteria, general filter type, and specific filter designs for cake filtration. It must be emphasized that for this type operation the filtration characteristics of the suspension depend for the most part on the resistance and porosity of the deposited cake, which can readily be established by laboratory measurements. The filter medium in the ideal case acts only as a support for the rapidly formed cake, which is the real filter base throughout most of the filtration cycle. Thus, provided a filter medium is used which allows cake formation without penetration and plugging of filter medium, the filtration performance depends only on the filter cake properties. In practice, selection of a filter medium which approximates this requirement is not easy, but usually can be achieved. The important point here is that in cake filtration the type filter which can be used is not usually limited by filter-media considerations, and laboratory measurements of cake properties can be interpreted in terms of filtration per-

formance on most filter designs with reasonable accuracy. This is not the case for solution clarification.

The initial selection between pressure and vacuum filtration for cake-filtration service can usually be made on the basis of the suspension alone if the process is continuous and continuous filtration is desirable. Where a batch process is involved or where capacity requirements are small, a batch pressure filter is generally to be preferred over a batch or continuous vacuum filter. In general a suspension can be handled satisfactorily by vacuum filtration if the specific cake resistance is less than 5×10^{11} ft./lb., or in more simple terms, if a 1/8-in. thick cake can be formed in 5 min. or less at 20-in. Hg. vacuum. With many highly flocculated materials of small ultimate particle size, pressure filtration may offer little advantage over vacuum filtration because of the high degree of cake compressibility; this becomes obvious when complete laboratory data are taken, and emphasizes the importance of taking such data before filter selection is made.

The majority of pressure filters for cake filtration is of the batch type from which the cake solids are discharged intermittently during periods of filter shutdown. The filter press is the simplest and most flexible design in this class. It is of particular value when dry-solids recovery is important, when succeeding batches of different materials must be handled without contamination, when the filter medium must be changed frequently, when it is desired to use wood as a material of construction, or when the process is new and batch size or character of slurry is likely to vary. The principal advantage of the filter press over other pressure filters for cake filtration are:

1. Adaptability to nearly all suspensions.
2. Simplicity of operation.
3. Low first cost and depreciation.
4. Availability in wide variety of designs and materials of construction.
5. Adaptability to high pressure operation up to 300 lb./sq.in.
6. Accessibility of filter medium.
7. Ability to produce dry washed cake.
8. Low power requirements and large filter area per unit floor space.

The chief disadvantages of this unit are:

1. High labor cost especially when handling large tonnages of solids.
2. High filter cloth costs because of mechanical wear.
3. Dependence of operating cycle on time necessary completely to fill frame, this requiring an excessively long time for highly compressible materials in order to get good washing and dry cake discharge.
4. Poor washing and sloppy cake discharge which result if frames are not filled and properly vented.
5. Leakage, cake exposure, and cake handling which may prevent use with volatile solvents or toxic materials.

A number of designs of pressure-leaf filters exists which are suitable for cake filtration. In general the leaves are mounted in the cylindrical filter tank either perpendicular or parallel to the cylinder axis, the tank may be orientated either horizontally or vertically, although leaves are always vertical. The leaves may be removed or exposed for dry cake discharge, or the cake may be removed by a sluicing spray and reslurried. In some designs the leaves can be rotated to facilitate uniform cake deposition or to aid in cake discharge by sluicing or by blowback and brushes or doctor blade. The pressure-leaf filter is less flexible than the press in cake-filtration service, but is particularly adapted (1) when solution recovery is the purpose and a sluicing discharge of cake solids can be used, (2) when volatile or toxic liquids are being handled, and (3) when close temperature control is necessary. Although certain pressure-leaf filters can be operated with dry-cake discharge, this is not recommended except under special conditions.

The principal advantages of pressure-leaf filters over other pressure filters in cake-filtration service are:

1. Reduced labor costs (sluicing discharge) over filter press.
2. Less mechanical wear of filter medium.
3. Efficient washing with some materials provided careful operating procedures are observed.

The chief disadvantages of this unit for cake filtration are:

1. High cost if operation is desired above 60 lb./sq.in.
2. Necessity for intelligent operation to prevent cake dropping and poor washing.
3. Lack of agitation except by recirculation, which, except at low solids concentrations, tends to result in sedimentation, nonuniform cakes, and poor washing.
4. Inability to control final cake thickness with feed of varying filterability or solids content, with consequent danger of bridging between leaves.
5. Inability to produce dry hard cake in most cases.
6. Considerable time and labor required to change filter medium.

The perforate basket centrifuge can be classified as a batch pressure filter and has a rather limited but important field of application. It is well suited to the handling of crystalline or granular materials in the 10-500 μ-size range as well as fibrous materials, and is particularly useful when effective washing and low cake moisture are desired. The perforate centrifuge is not suited to the majority of chemical filter cakes because of their compressible nature. Disadvantages of perforate basket centrifuges are relatively high initial and maintenance costs, as well as high labor costs.

The horizontal plate filter in cake-filtration service combines several advantages of filter press and pressure-leaf filters, but is limited to relatively small sizes (under 150 sq.ft.), and remains a high labor cost filter. This filter design is best suited for cake filtration where dry recovery of valuable cake is desired, capacity requirements are small, and total enclosure is dictated by volatile or toxic solution.

The principal advantages of horizontal plate filters in cake filtration service are:

1. Totally enclosed design with close temperature control possible.
2. Simplicity of operation.
3. Ability to obtain good washing and dry cake without complete filling of cake chamber.
4. Ability to filter all solution from filter casing before opening (scavenger plate).
5. Accessibility through interchangeable plate assemblies.

The chief disadvantages of these units for cake filtration are:

1. Higher investment cost.
2. Limited to 60 lb./sq.in. maximum operating pressure in standard designs.
3. Relatively high labor cost for dry cake discharge.
4. Limited capacity per unit.

Although a continuous pressure filter for cake filtration would seem desirable for handling slow filtering suspensions in continuous processes, or where labor for batch pressure filtration is high, the difficulties in design and high cost of continuous pressure filters have severely limited this development and usage. A possible exception is the continuous screen centrifuge which is here classified as a continuous pressure filter and has been rather widely applied to the handling of granular crystals larger than 150-mesh.

Continuous pressure filters have been built both as continuous rotary-drum and continuous rotary-disk units, internal construction resembling that of similar-type multiple-compartment vacuum filters. The drum type has been built in drum areas to 100 sq.ft. for 30 lb./sq.in. working pressure, and in larger areas for lower working pressures. Disk-type units have been built for 100 lb./sq.in. service and in areas up to 600 sq.ft. Continuous automatic pressure filters of these types are specially suitable when volatile solvents are used, when the pressure makes continuous operation possible or gives a sufficiently dryer cake than a continuous vacuum filter to warrant the greater investment, and when filtering materials at temperatures which are not ap-

plicable to continuous vacuum filtration. The major determent to wider usage of continuous pressure filters of these types is the high investment cost and elaborate controls necessary. In most cases these costs cannot be justified by labor savings over batch operation.

The simplest form of commercial vacuum filter is the vacuum Nutsche which is simply a false bottom tank with vacuum applied below the false bottom. Such batch vacuum filters are particularly suited when free filtering crystalline materials are being recovered, when severely corrosive conditions require use of ceramics, and when small quantities of a number of different valuable cakes must be handled. The advantages of this batch vacuum filter over other filters in cake filtration service are:

1. Simplicity of construction and operation.
2. Effective displacement wash.
3. Minimum cloth plugging.
4. Versatility as to type material handled.

Disadvantages of this type filter are:

1. Difficulty and high labor cost of cake removal.
2. Large floor space per unit area (limited to about 100 sq.ft. in most cases).

The vacuum-leaf filter is similar to the pressure-leaf filter in the principle of batch operation, but because of its open construction is simpler to operate and is capable of better washing than pressure-leaf filters. Vacuum-leaf filters are best considered for use when large tonnages of solids are to be filtered and when especially effective washing is necessary with maximum recovery of strong filtrate.

Advantages of this unit over other filters in cake-type service are:

1. Simplicity of operation and ease of control.
2. Low first cost per unit area when filter area is very large, $>$ 1000 sq.ft.
3. Low labor cost when cake discharges freely.
4. Easy inspection and replacement of cloths or leaves.
5. Ease of construction in corrosion-resistant materials.

The chief disadvantages of this type batch vacuum filter are:

1. Large size of building required.
2. Short cloth life when blowback and hand scraping are used for cake discharge.
3. Limitations on liquid volatility or temperature imposed by open construction and vacuum operation.

In contrast to continuous pressure filters, continuous vacuum filters are made in many designs and are widely used for cake filtration. They are best adopted when the process is either continuous or else batchwise in fairly regular sequence of batches. Comparative investment is high for small capacity but drops rapidly for larger capacity while labor savings increase rapidly. In general their application is limited only by the filterability of the feed, and by the filtration temperature and altitude which limit the vacuum without flashing.

The multiple-compartment drum filter is the most widely used type and is available in many special forms, all of which involve sectionalized surface compartments, internal filtrate piping, and a rotary valve. This type continuous filter is suited to dry-cake recovery, solution recovery and cake reslurrying, or any combination of cake washing, solution recovery and dry-cake recovery. Advantages of this type vacuum filter in cake filtration service are:

1. Operation is flexible and operating labor is low.
2. Can be made of most materials of construction including wood.
3. With aid of auxiliary devices, effective cake washing and dewatering can be obtained in most cases.
4. With special construction, cakes as thin as $\frac{1}{8}$ in. can be handled satisfactorily.
5. By special design this continuous filter is adaptable to a wide variety of suspensions and purposes. The top-feed-type adapted for the filtering and drying of coarse granular solids is a good example of one of several special designs.

The major disadvantages of this type continuous vacuum filter are:

1. Relatively high cost in small sizes.
2. Limitation imposed on displacement wash effectiveness and strong filtrate recovery by internal piping holdup.
3. Pressure drop within internal piping of drum may limit capacity.
4. Cloth replacement cost may be high. Cloth life may be short because of mechanical wear if doctor blade is used, or may be limited by blinding even if special discharge mechanisms are used, since the short operating cycle promotes both cloth wear and blinding.

The single-compartment-type continuous drum filter has no internal drum piping or rotary valve avoiding some of the operating difficulties inherent in the multiple compartment design; however, evacuation of the entire drum, as well as close construction tolerances required for efficient blowback discharge of the cake, make construction costs higher per unit area. This continuous vacuum filter is best suited for use when handling slow filtering suspensions requiring thin cake discharge, and when an effective displacement wash and strong filtrate recovery are desired. Special designs are also well suited to fast filtering suspensions such as pulp fibers and granular polymers. Advantages of the single-compartment filter over the multiple-compartment filter are:

1. Ability to discharge successfully cakes less than $\frac{1}{16}$-in. thick enables continuous filtration of more difficult filterable material.
2. Absence of internal piping and rotary valve reduce filtrate holdup and mixing, thus improving effectiveness of displacement wash.
3. Ability to operate with thin cakes and absence of internal pressure drop result in increased capacity per unit area.

Disadvantages of single-compartment construction compared with multiple-compartment construction are:

1. Less flexible operation because of fixed submergence and the single operating vacuum available for all portions of operating cycle.
2. Cost per unit area is higher because of rugged construction required. Also wood cannot be used as a material of construction.
3. Vacuum capacity requirements are appreciably higher because of air leakage around blowback shoe.
4. Maximum size is limited to 140 sq.ft./filter, because drum must withstand atmospheric pressure.

The design of the rotary-disk-type continuous vacuum filter makes practical the construction of filters with large filtration areas. This type continuous vacuum filter is best suited when handling suspension of high solids content, when a relatively thick and easily dischargeable cake is formed, and when the cake need not be washed. The chief advantages of the disk-type filter over drum-type continuous vacuum filters are:

1. The much larger filter area per unit floor area and lower cost per unit area.
2. Sector construction permits filter media replacement by sectors as required.
3. Design can be modified to handle two different feed suspenions simultaneously on a single unit.

Disadvantages are:

1. Inability to wash effectively.
2. Poorer dewatering and inability to use auxiliary dewatering devices.
3. Inability to handle suspensions tending to settle rapidly.
4. Cake sloughing is more pronounced and cake cracking tends to cause cake to drop off vertical leaves.
5. Poor performance and short cloth life when handling cakes which smear on discharge or require strong blowback.

Continuous vacuum filters of the horizontal type are useful in handling free-filtering materials where filtration rates are high and solids tend to settle rapidly.

Maximum filtration area is limited to 125 sq.ft. or less, floor space requirements are large, and cost per unit filtration area is higher than for drum or disk filters. They are well suited to cake filtrations where exceptionally good washing is required since the cake is completely covered with wash liquor and no distribution problem exists. The horizontal rotary filter is of multiple-compartment design with rotary valve for splitting filtrate fractions. The major advantages of the horizontal rotary design are:

1. Settling of feed aids filtration, and material of wide particle size can be handled.
2. Ability to employ effectively countercurrent washing.
3. Construction is simpler than a drum filter and operation is more flexible.

Disadvantages are:

1. Larger floor space requirement and slightly higher unit cost than drum filter.
2. Discharge method results in cake heel remaining and makes impractical the handling of any but free filtering materials.

The design of the horizontal band filter eliminates the multiple-compartment design by passing the filtering surface over fixed suction boxes for collection of various filtrate cuts; in this, its principle resembles the single-compartment drum filter. Advantages of horizontal band filter are:

1. Settling aids filtration as with horizontal disk filter.
2. Ability to obtain very complete displacement wash and strong filtrates recovery because holdup and mixing in internal piping are eliminated.
3. Flexibility and ease of countercurrent wash.
4. Ability to handle moderately slow filtering materials because of complete cake discharge and filter medium rinse between cycles.

The chief disadvantages are:

1. High initial cost and large floor space requirements.
2. Limitation in corrosive service imposed by use of rubber supporting band.
3. Requirement that filtrate wet rubber and metal to form vacuum seal.

Solution Clarification

In Figure 3 the general selection criteria, general filter types, and specific filter designs for solution clarification are shown. It must be emphasized that for this type operation the filtration characteristics of the solution are entirely dependent on the filter medium or bed used, and that prediction of filtration performance can be made only on the basis of the same filter medium or bed.

The specific filter designs may then be logically classified as those using filter media or cartridges, those using a filter-aid bed, and those employing a granular bed as the active filter element.

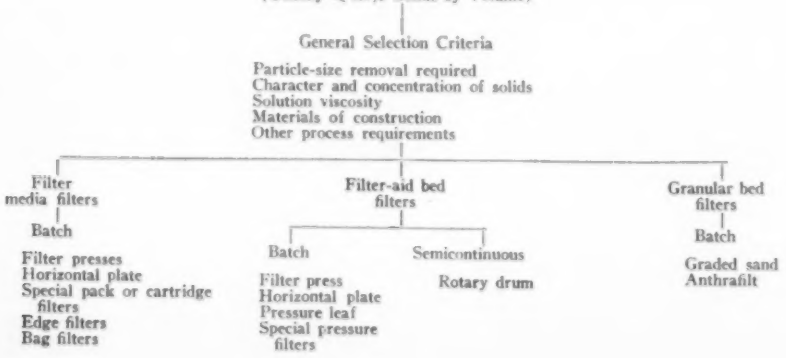
Units employing a filter medium or cartridge as the active filtration element are best adopted where solids concentration is below 50 p.p.m. since this permits long cycles and economical filter-media life. Likewise they are well adapted to small throughput rates as this minimizes filter-media costs. The filter medium can be provided to give any desired clarity to the extreme of biological filtration, and the filter medium may be highly purified to prevent contamination of solution. Thus, this class of filters is ideally suited to polishing operations where the volume of solution to be handled is not large and the solids concentration is low. For large throughput rates or solids concentrations above 50 p.p.m. the filter-media costs may become excessive.

The filter press is the most flexible unit in this class. With asbestos sheets, nonwoven fabrics, or filter paper it is suited to the polishing of many low-viscosity liquids. With special filter-media packs of these same media or felt, canton flannel, filter cotton, or Kimpak wadding, the press is suited to the clarification of high viscosity solutions up to 1000 poises or higher, and can be designed for filtration pressures up to 300 lb./sq.in. in such service. As with most of the filters in this class, the filter medium and labor costs become excessive except at very low solids content. As with the filter press in cake filtration, leakage may be a limiting factor if toxic or volatile liquids are handled.

The horizontal plate filter may be used in clarification service with asbestos sheets, filter paper, etc., but is not well suited to packs of several filter media. A wide range of clarity can be obtained with these media and liquids of viscosity up to 50 poises can be handled satisfactorily. The major advantage of this design over the filter press, when using sheet media for clarification, is the enclosed construction which enables handling volatile liquids or material requiring close temperature control.

A wide variety of small pack or cartridge filters is available for solution clarification. The designs vary and each uses single or multiple-filter elements of special design. These are usually of graded density to minimize the plugging rate and are available in grades which vary in particle-size removal characteristics from 50μ removal down to removal of particles < 1-μ diam. These units are ideally suited to the polishing of flows < 20 gal./min. and are relatively cheap in this type service. With solids contents as high as 100 p.p.m. and flows in the range of 200 gal./min., the cost of filter cartridges becomes about twice the cost of filter aid for clarifica-

Fig. 3. Solution clarification. (Usually $< 0.1\%$ Solids by Volume)



tion through filter-aid beds. These pack or cartridge filters are not suitable for operation at viscosities above 10 to 15 poises and operating pressure is limited to a maximum of from 60 to 100 lb./sq.in.

Edge filters are made only in small sizes suitable for polishing rather small flows. Elements of ring packs or ribbon packs are available for particle-size removal form 0.5-μ to 50-μ size. The ring-pack elements can be readily cleaned by disassembling or backwashing and are used widely for fuel clarification.

Bag or hat filters are the simplest form of clarification filter, employing the filter medium in the form of a bag and operating by gravity. The bag often consists of a multiwalled bag of supporting fabric and filter cotton, and may be simply strapped over a pipe end. Simplicity is the chief advantage, while high filter-media costs and limitation to low-pressure operation are the major disadvantages.

Clarification filters employing a filter-aid cake find wide application with solution having solids contents from only a few p.p.m. to 0.1% by volume, or greater. They are particularly adapted to large throughputs at low-solids concentration or to the handling of solids concentrations above 100 p.p.m. Generally both a filter-aid precoat and the addition of filter aid to the feed suspension are required, although one or the other can sometimes be eliminated. This requires a more elaborate feeding and operating technique than with the filter-media filters. A wide variety of diatomaceous earth and fibrous filter aids are available, making an optimum balance between clarity and flow rate possible in most cases. The filter aids themselves place several limitations on this type of operation, however. Most commercial diatomaceous earth filter aids contain considerable mineral impurities which may contaminate the solution. Purified diatomaceous earth, and pure asbestos or cellulose fibers are available at higher cost. Practical operating viscosity is limited to about 50 poises with the fastest diatomaceous earth while some fibrous materials are useful from 200-300 poises. Where high clarity is desired, filter-aid filters must be followed often by a polishing filter of the filter-media type to prevent some filter-aid contamination. Thus, filters clarifying by use of a filter-aid bed are best suited to applications requiring high throughputs or where solids content is relatively high, and in general do not give the degree of polishing possible with a filter-media filter.

The filter press and horizontal plate filters, as used for solution clarification with filter media only, are equally well adapted to use with filter aids. In this case the horizontal plate filter has the added advantage that a uniform precoat is more easily established and intermittent operation is possible without having difficulty with dropping of filter-aid cake and poor clarity.

The pressure-leaf filter, which is not suited to solution clarification with filter media alone because of the difficulty of filter media changes, is well suited to and widely used with a filter-aid bed. The chief advantages are the closed construction, permitting use with volatile liquids, the large sizes which are practical, and the automatic sluicing or brushing discharge available with most models. The major disadvantages are the usual pressure limit from 60 to 100 lb./sq.in., and the leaf construction which limits practical operating viscosities to 20 poises or less.

A number of special design filters are available for use with filter-aid beds. These employ special filter elements of porous stone or carbon which permit operation at high temperatures under corrosive conditions, or special wire wound or porous stainless steel elements which permit operation at high pressure drops. Their principal applications stem from these features and their higher cost usually requires justification on the basis of such conditions.

The rotary-drum filter is the only form of semicontinuous filter for solution clarification, and may be operated under vacuum or pressure to 30 lb./sq.in. A thick precoat layer is initially deposited and cut away a few thousandths of an inch during each revolution. This type filter is best suited to the clarification of solutions of relatively high solids content (0.1% and higher), since the higher cost of continuous solids discharge is then more easily justified. For this condition average filtration rates are appreciably greater than for batch filters and labor is greatly reduced. At lower solids content this advantage decreases, while penetration of the precoat layer tends to be greater requiring a greater precoat cut, thus shortening over-all precoat life. Operation below 100 p.p.m. solids content is usually not practicable. Gradual cutting away of precoat surface limits the filter aid which can be used to diatomaceous earth in most cases.

Rotary-drum filters of the single-compartment type have been used for continuous precoat clarification in which fresh precoat is deposited and discharged from the drum during each revolution. Such operation is generally impractical because of high filter-aid costs, unless a cheap available material such as paper pulp or sawdust is used.

Granular-bed filters for solution clarification are associated with water filtration, and are suited to handling large volumes of solution where solids can be effectively removed by backwashing. In general the usual granular-bed filter will not remove fine solids from low-viscosity solution unless they are highly flocculated and this limits process applications. Also, the volume of backwash required is a serious drawback with most process solutions unless a water backwash can be used; this involves dilution of process solutions and is often expensive. Low initial and operating costs, simplicity of operation, and general corrosion resistance are principal advantages of such filters in their usual form. They generally use graded sand or Anthrafilt in the size range of 0.2 to 0.8 mm.

Granular-bed filters of finer bed particles can also be used effectively for clarification of small quantities of high viscosity solutions. In this case the pressure drop may be several hundred pounds per square inch and the mechanism of homogenization may accompany filtration.

Summary

In summary the question of "What Type Filter and When," is not so simple that it can be answered here in a specific manner. Final filter selection is usually not a cut-and-dried procedure, but rather involves detailed paper evaluations as well as extensive experimental work with the process material. In this respect filters are no different from most other forms of chemical processing equipment.

Literature Cited

1. Miller, S. A., *Chem. Eng. Progress*, **47**, No. 10, 497 (1951).

CORRECTION

In "Temperature Gradients in Gas Streams Flowing Through Fixed Granular Beds," authored by C. A. Coberly and W. R. Marshall, Jr. ("C.E.P." March, 1951) the following correction is made:

On page 144, Table 1, column 2, the values of G should be rearranged as follows:

TABLE 1.—EXPERIMENTALLY DETERMINED GAS TEMPERATURES AT VARIOUS RADIAL POSITIONS AND HEIGHTS IN A 5-IN. DIAMETER PACKED TUBE

Run No.	G, lb./(hr.)(sq.ft.)
1—1 through 1—6	1215
2—1 through 2—6	1005
3—1 through 3—6	750
4—1 through 4—6	475

PASSAGE OF SOLIDS THROUGH ROTARY KILNS

Factors Affecting Time of Passage

W. C. SAEMAN*

Equations based on theoretical considerations have been developed to give time of passage, axial transport velocity, bed depth, and other characteristics related to the passage of material through rotary kilns without flights. The equations are corroborated by experimental data, and are particularly useful for evaluating the effects of constrictions in kilns and to calculate the bed depth in Roto-Louvre driers. This work fills a gap in the fundamental knowledge of kiln design and operation, and the theoretical equations may be used to supplant empirical methods of calculating kiln performance.

A RATHER extensive experimental study of the relation of time of passage of solids through rotary kilns to the measurable kiln variables was reported by Sullivan, Maier, and Ralston (*4*). The correlations that these authors made of their data are empirical, as are the correlations later proposed by other authors (*1, 2*).

The factors which affect the time of passage of material through a rotary kiln are quite obvious and have been identified in the references cited above. These factors were related by simple, fundamental relations known to exist in rotary kilns to yield a theoretical analysis of this problem. This analysis, which is the subject of the present paper, resulted in development of an equation for time of passage through lightly loaded kilns that closely resembles the basic empirical one of Sullivan and coworkers, and also in development of equations that adequately express data for kilns with relatively heavy loads. The theoretically derived equations clarify several previously obscure characteristics of kiln operation, such as the effects of feed rate and constrictions on the time of passage, and are applicable to calculation of time of passage through Roto-Louvre type driers, which have tapered diameters.

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Geometric Analysis of Factors

Development of Transport Velocity Equation. The path that a particle follows in traveling through a rotary kiln was described in detail by Sullivan, et al. (*4*). These investigators found that the position of a particle in the bed of material in a kiln remains stationary with respect to the kiln until it reaches a position close to the surface of the bed, and that from this point it cascades down along the bed surface in such a direction as to progress a short distance axially through the kiln and comes to rest at some lower position; the cycle is then repeated. The paths a particle may follow, based on this observation, are illustrated by Figure 1. Each possible path may be designated by the radius of the path and, as will be shown, the rate at which a particle progresses through the kiln (axial transport velocity) is dependent on the radius of the path followed and on several measurable quantities that are characteristic of the kiln and of the material. A particle generally does not follow the same path on successive cascades but, in the considerations that will be made of the average behavior of all particles in the kiln, the variations in paths of individual particles will be of no significance since all the possible paths are obviously followed constantly by the same number of particles.

The axial transport distance of a given particle per cascade may be related to the path radius and to measurable variables as shown in the geometric construction in Figure 2. This construction represents a section of a rotating kiln between planes that cut the kiln axis perpendicularly through points *O*

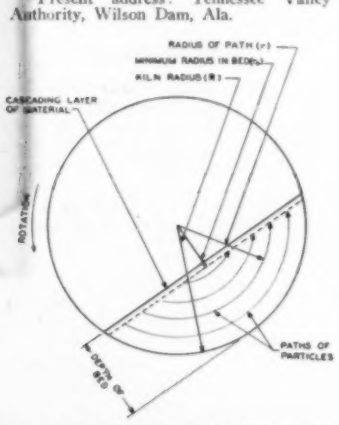

Fig. 1. Cross section of rotary kiln showing paths followed by particles.

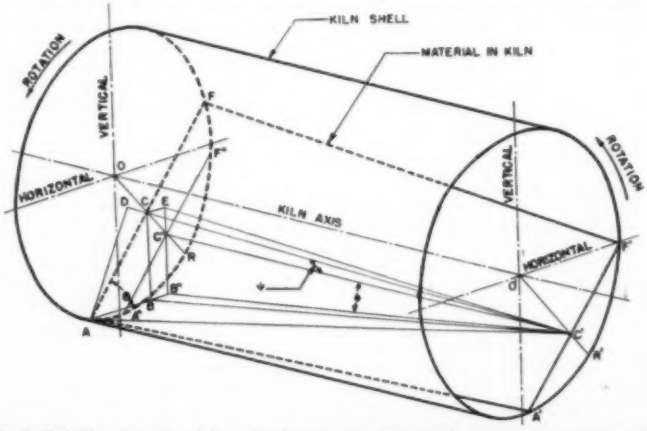

Fig. 2. Projection of a rotary kiln section showing geometry for derivation of axial transport rate of a particle.

and O'. The kiln slopes downward from left to right, and the bed depth decreases in the same direction. Line AF is the intersection of the bed surface with plane O, and $A'F'$ is the intersection of that surface with plane O'. Points C and C' are the midpoints of AF and $A'F$, respectively. The length of the kiln section shown is such that line AC', which lies in the bed surface, is a horizontal line. The derivations that are to follow are based on the postulate that particles cascading down the bed surface under the action of gravity will cascade in a direction that is perpendicular to this horizontal line AC'. The particle that will be considered is one that cascades along line DA, which was constructed in the plane of the bed surface and perpendicular to AC'. Point D, which is on CC' extended, represents the midpoint of this particle's cascade if it is assumed that the particle cascades from kiln wall to kiln wall. The path of this particle may be designated by its radius OR, which is the kiln radius. It is pointed out, however, that similar geometric constructions apply to particles with paths of any radius, r; a particle with a path corresponding to the kiln radius was chosen for the present consideration only to simplify Figure 2.

One half of the axial transport distance of the particle per cascade is represented by DC. From trigonometry we have

$$DC = \frac{(AC)^2}{CC'} \qquad (1)$$

The length CC', is not a directly measurable quantity; however, it may be related to measurable quantities by appropriate methods as shown below.

In Figure 2, line $A'F'$ was projected onto plane O as line $A''F''$. Point C'' is the midpoint of $A''F''$ and, consequently, coincides with projection of point C' onto plane O, and line $C'C''$ is parallel to the kiln axis. Line AB'' is the intersection of plane O with a horizontal plane through the previously described horizontal line AC'. The angle between this horizontal plane and the surface of the material is the angle of repose of the material as defined (*4*). In the figure, this angle is equal to the angle between line AD and the horizontal plane and is approximately equal to θ, which is the angle CAB''. Point B'' lies in a vertical plane through line $C'C''$ thus, angle $C''C'B''$ represents the angle of the kiln axis with the horizontal and is equal to the slope of the kiln. The angle $CC'C''$ is the angle between the surface of the bed and the kiln axis measured in a plane perpendicular to the surface of the bed.

Returning now to the problem of relating CC' to measurable variables, passage of a vertical plane through line CC' gives intersections that form triangle $CC'B$ and from the law of sines,

$$CC' = CB\frac{\sin CBC'}{\sin CC'B} \qquad (2)$$

Further consideration of the figure shows that $CB = AC \sin \theta$. Also, since angle CBC' is approximately a right angle,

$$\sin CBC' \doteq 1^*$$

Thus, Equation (2) may be rewritten

$$CC' = AC\frac{\sin \theta}{\sin CC'B} \qquad (3)$$

Since angle $CC'B$ is the same order of magnitude as the kiln slope, and is therefore generally small,

$$\sin CC'B \doteq \text{angle } CC'B$$

Projection of triangle $CC'B$ onto the vertical plane through $C'C''$ gives triangle $EC'B''$ which, since angle $BC'B''$ is generally small, is approximately similar to tiangle $CC'B$. Thus,

$$\sin CC'B \doteq \text{angle } EC'B''$$

and Equation (3) may be rewritten

$$CC' = AC\frac{\sin \theta}{\text{angle } EC'B''} \qquad (4)$$

From the construction in Figure 2, angle $CC''E = \theta$, thus,

$$\cos \theta = \cos CC''E = \frac{\tan EC'C''}{\tan CC'C''}$$

Since angles $EC'C''$ and $CC'C''$ are generally small,

$$\tan EC'C'' \doteq \text{angle } EC'C''$$

and,

$$\tan CC'C'' \doteq CC'C''$$

thus,

$$\text{angle } EC'C'' \doteq \cos \theta \text{ angle } CC'C'' \qquad (5)$$

The angle $EC'C''$ in Equation (5) plus angle $C''C'B''$ is equal to angle $EC'B''$ in Equation (4). The latter equation may therefore be rewritten as

$$CC' = \frac{AC \sin \theta}{\text{angle } C''C'B'' + \cos \theta \text{ angle } CC'C''} \qquad (6)$$

Substitution of this expression for CC' in Equation (1) gives,

$$DC = \frac{AC(\text{angle } C''C'B'' + \cos \theta \text{ angle } CC'C'')}{\sin \theta}$$

Substitution of simplified symbols in this equation gives the following expression for the axial transport distance per cascade:

$$x_c = \frac{c(\phi + \psi \cos \theta)}{\sin \theta} \qquad (7)$$

where

x_c = axial transport distance of particle per cascade ($x_c = 2DC$)
c = length of chord at surface of bed ($c = 2AC$)
ϕ = slope of kiln (angle $C''C'B''$), radians
ψ = angle between surface of material and kiln axis (angle $CC'C''$), radians
θ = angle of repose of material (approximately)

The average transport velocity of a particle for a specific path in the kiln is equal to the product of the displacement of the particle for a single cascade and the number of cascades per unit time. The number of cascades per unit time is equal to the product of the speed in revolutions per unit time and the number of cascades per revolution. The number of cascades per revolution may be expressed as:

$$N = \frac{\pi}{\sin^{-1}\frac{c}{2r}} \qquad (8)$$

where

N = number of cascades per revolution
r = radius of particle path in bed (see Fig. 1)

The product of Equation (7), Equation (8), and the kiln speed gives the average transport velocity; thus,

$$u = nc\left(\frac{\phi + \psi \cos \theta}{\sin \theta}\right)\left(\frac{\pi}{\sin^{-1}\frac{c}{2r}}\right) \qquad (9)$$

where

u = average axial transport velocity of a single particle
n = speed of kiln

Application of Equation to Calculation of Time of Passage Through Lightly Loaded Kilns. For lightly loaded kilns, the angle

$$\sin^{-1}\frac{c}{2r}$$

is approximately equal to the ratio $c/2r$. Using this approximation in Equation (9) gives,

$$u = 2\pi rn\left(\frac{\phi + \psi \cos \theta}{\sin \theta}\right) \qquad (10)$$

The length of the kiln divided by u gives the time of passage,

$$t = \frac{L \sin \theta}{2\pi rn(\phi + \psi \cos \theta)} \qquad (11)$$

where

t = average time of passage of a single particle
L = length of kiln

In actual practice, lightly loaded kilns usually operate with essentially uniform bed depth along their length, and effects of nonuniformity such as result from

* The symbol $\doteq$ indicates an approximate equality.

droop at the discharge end of the kiln are small and can be neglected. For cases of this kind, the angle ψ in Equation (11) is zero and the time of passage of a particle following a path of radius r is

$$t = \frac{L \sin \theta}{2\pi r n \phi} \qquad (12)$$

For the lightly loaded kiln, the effective average radius of the particle paths closely approaches the kiln radius R, which may therefore be substituted for r in Equation (12). With such a substitution, Equation (12) corresponds closely to the basic empirical formula of Sullivan et al. (*4*) for lightly loaded kilns; their equation was

$$t = \frac{1.77L(\theta^\circ)^{1/2}}{2Rn\phi^\circ} \qquad (13)$$

where

t = time of passage through kiln, min.
L = length of kiln, ft.
θ° = angle of repose of material, deg.
R = kiln radius, ft.
n = speed of kiln, rev./min.
ϕ° = slope of kiln, deg.

The difference in the functional form of θ in Equations (12) and (13) does not have a pronounced effect on calculated values of the time of passage, since θ normally varies between relatively narrow limits (angle of repose usually varies between 35° and 45°).

The similarity of Equations (12) and (13) may be shown by consideration of the case where $\theta = 40°$. The time of passage based on the empirical equation is

$$t = \frac{11.2L}{2Rn\phi^\circ}$$

This compares closely to the time of passage calculated from the theoretically derived equation, which is

$$t = \frac{11.7L}{2Rn\phi^\circ}$$

Since the basic empirical formula of Sullivan, et al. summarizes the bulk of their experimental data for lightly loaded kilns, the agreement which is seen to exist between the empirical and theoretical relations shown above serves as experimental confirmation of the validity of the theoretical analysis that has been presented thus far in the present paper.

Extension of Relations for a Single Particle to a Volumetric Rate Basis

Development of General Equation. For relatively heavy kiln loads the approximation

$$\sin^{-1}\frac{c}{2r} = \frac{c}{2r}$$

used in Equation (10) is no longer valid. Equation (9) may, however, be extended, as described below, to a form that is applicable to rotary kilns with heavy loads and with variable bed depth.

Equation (9) expresses the rate of transport of a particle as a function of the radius of the path of the particle. This equation can be transformed to a volumetric basis by multiplying each side by the increment of area associated with the path r. The increment of area is

$$dA = 2r \sin^{-1}\frac{c}{2r}\, dr \qquad (14)$$

Multiplying Equation (9) by Equation (14) gives

$$ud\mathcal{A} = \pi c\left(\frac{\phi + \psi \cos\theta}{\sin\theta}\right)\left(\frac{\pi}{\sin^{-1}\frac{c}{2r}}\right)\left(2r \sin^{-1}\frac{c}{2r}\right) dr$$

which simplifies to

$$ud\mathcal{A} = 2\pi r n c\left(\frac{\phi + \psi \cos\theta}{\sin\theta}\right) dr \qquad (15)$$

The length of the chord, c, may be expressed as a function of the path radius by

$$c = 2(r^2 - r_o^2)^{1/2} \qquad (16)$$

where r_o is the path in the bed of material having the shortest radius (see Fig. 1). Substituting Equation (16) in Equation (15) gives

$$ud\mathcal{A} = 4\pi n\left(\frac{\phi + \psi \cos\theta}{\sin\theta}\right)(r^2 - r_o^2)^{1/2} r\, dr \qquad (17)$$

Since the transport velocity and weight of material associated with each path remain unchanged whether or not the material follows the same path consistently on successive cascades, the average volumetric transport rate, q, may be obtained by direct integration of Equation (17) from $r = r_o$ to $r = R$, or

$$q = \frac{4\pi n}{3}\left(\frac{\phi + \psi \cos\theta}{\sin\theta}\right)(R^2 - r_o^2)^{3/2} \qquad (18)$$

where q = volumetric transport rate in kiln.

It should be noted that the shortest radius path, r_o, does not coincide exactly with the surface of the material but differs from it by the thickness of the cascading layer of material which, however, is usually negligible.

The volumetric transport rate, q, is related to the average transport velocity at a given point in the kiln by the following equation

$$u_{avg.} = \frac{q}{A} \qquad (19)$$

where A = cross-sectional area of the bed.

For a given r_o, either measured or calculated from Equation (18), the cross-sectional area of the bed may be calculated or may be obtained from handbook tabulations (*3*). If the depth of the bed is essentially constant along the length of the kiln, the bed cross section is constant and the average time of passage of material through the kiln is

$$t = \frac{LA}{q} \qquad (20)$$

If, however, the cross-sectional area does not remain constant along the length of the kiln, Equation (20) must be written in differential form,

$$dt = \frac{Adx}{q} \qquad (21)$$

from which the total time of passage can be obtained by integration.

Determination of Maximum Kiln Capacity. The volumetric transport rate at any point in a kiln as a function of r_o at that point is given by Equation (18). Differentiation of this equation with respect to r_o gives

$$\frac{dq}{dr_o} = -4\pi n\left(\frac{\phi + \psi \cos\theta}{\sin\theta}\right)(R^2 - r_o^2)^{1/2} r_o \qquad (22)$$

If dq/dr_o is equated to zero, the resulting equation is satisfied by $r_o = R$ and $r_o = 0$. It is obvious that the solution $r_o = R$ corresponds to the minimum volumetric transport rate, while $r_o = 0$ corresponds to the maximum transport rate. This result indicates that the maximum volumetric transport rate through a kiln is obtained by operating with the bed depth in the kiln equal to the kiln radius.

Time of Passage Where Bed Depth Is Uniform. Where the depth of the bed in a kiln is uniform along the kiln length, ψ in Equation (18) is zero, r_o remains constant, and the equation can be solved directly for any of the factors involved. Experimental corroboration of Equation (18) for uniform bed depth is provided by data for the uniformly loaded section of a kiln reported by Sullivan, et al. (*4*). This corroboration is discussed below.

Bed depths measured by the above-mentioned authors in the uniformly loaded section of an experimental kiln under various operating conditions are tabulated in Table 1, together with values calculated for the same conditions by use of Equation (18).

It may be seen that for the 267 cc./min. feed rate there is close agreement between the measured bed depths and those calculated from Equation (18). For the relatively low feed rate of 25 cc./min., the bed depths calculated by use of Equation (18) are significantly smaller in all cases than the depths measured directly by Sullivan et al. However, the former are in good agreement with the depths (see Table 1) calculated from time of passage and feed

rates measured by those authors, which indicates that time of passage calculated by use of Equation (18) in conjunction with Equation (20) would agree closely with measured time of passage.

Time of Passage Where Bed Depth Is Not Uniform. Nonuniform depth of material occurs in a kiln near the discharge end and above constrictions. In cases where the bed is relatively deep, reliable calculations of time of passage by use of Equation (18) can be made only by substitution of appropriate values for ψ, the slope of the bed with respect to the kiln axis, along the entire kiln length. For such calculations, ψ can be expressed in terms of the variation of r_o along the length of the kiln; thus,

$$\psi = \frac{dr_o}{dx}$$

Substitution of dr_o/dx for ψ in Equation (18) gives the differential equation

$$\frac{dr_o}{dx} = \frac{3q \sin\theta}{4\pi n \cos\theta\,(R^2 - r_o^2)^{3/2}} - \frac{\phi}{\cos\theta} \quad (23)$$

Although the variables in this equation are separable, the formal integral solution of the equation is difficult. For the particular case where $\phi = 0$ the variables in Equation (23) reduce to the following standard form for integration

$$\int (R^2 - r^2)^{3/2} dr =$$

$$\left[\frac{r}{4}(R^2 - r^2)^{3/2} + \frac{3R^2 r}{8}(R^2 - r^2)^{1/2} + \frac{3R^4}{8}\sin^{-1}\frac{r}{R} = f(r)\right.$$

Using this integral in Equation (23) gives

$$\frac{4\pi n \cos\theta}{3q \sin\theta} f(r) = x + C \quad (24)$$

where C is the constant of integration. Equation (24) can be manipulated to yield a solution involving the slope, ϕ, but the work becomes more tedious than the solution to this problem warrants, since the applicability of Equation (23) itself becomes more and more questionable as the value of dr_o/dx becomes large due to the approximations involved in the derivation. At the discharge end of a kiln where droop in the bed reaches its maximum value, it is obvious that the inclination of the surface relative to the horizontal will not exceed the angle of repose of the material. This limitation establishes the contour of the surface of the material in the kiln independently of Equation (23). In the final solution the values of r_o from Equation (23) should merge smoothly with the limiting values at the end of the kiln, which is equal to the angle of repose reduced by an amount equal to the slope of the kiln ($\phi = \theta - \phi$). This can be conveniently done by graphical methods. Other approximate solutions can readily be obtained by series approximation, or by numerical integration using appropriate interpolation formulas.

The validity of Equation (23) was checked against the data of Sullivan, et al. for cases where they incurred large droop effects in their kiln loading. In making this check, the interpolation formula method was used for integrating Equation (23) as follows. Equation (23) was represented in general form as

$$\frac{dr_o}{dx} = \frac{1}{F(r_o)} \quad (25)$$

where $F(r_o)$ is an appropriate interpolation function so that $\int F(r_o)dr_o$ can be readily evaluated. Appropriate interpolation functions, $F(r_o)$, were then obtained by plotting values of the reciprocal of dr_o/dx, calculated from Equation (23), vs. r_o as is shown in Figure 3, drawing curves through these plotted points, and then approximating these curves by arbitrary algebraic functions which could be easily integrated. The specific functions with the limits of validity for the function used for a 1°-kiln slope were:

TABLE 1.—MEASURED AND CALCULATED VALUES FOR UNIFORM BED DEPTH IN AN EXPERIMENTAL KILN AT VARIOUS SLOPES *

	Depth of Bed, mm.				
	Nominal Feed Rate, 267 cc./min.		Nominal Feed Rate, 25 cc./min.		
Kiln Slope, Degrees	Measured †	Calculated from Equation (18)	Measured †	Calculated from Measured Time of Passage and Feed Rate	Calculated from Equation (18)
1	....	....	14.7	11.7	13.2
2	....	....	11.3	7.5	8.0
3	34.6	37.4	10.3	5.7	6.0
4	28.6	28.8	7.6	4.6	5.0
5	25.4	24.6	....	....	....
6	22.9	21.3	....	....	....

* Experimental data of Sullivan, et al. (4). Kiln was 7 ft. long by 76 mm. radius and was rotated 2.5 rev./min. Angle of repose of material was 35°.
† Depth measured with kiln in motion.

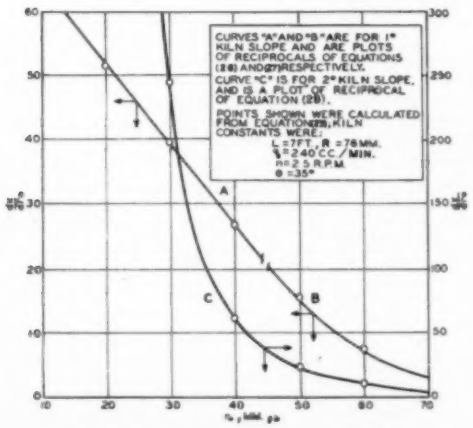
CURVES "A" AND "B" ARE FOR 1° KILN SLOPE AND ARE PLOTS OF RECIPROCALS OF EQUATIONS (20) AND (23) RESPECTIVELY.
CURVE "C" IS FOR 2° KILN SLOPE, AND IS A PLOT OF RECIPROCAL OF EQUATION (20).
POINTS SHOWN WERE CALCULATED FROM EQUATION 23, KILN CONSTANTS WERE:
L = 7 FT., R = 76 MM.
q = 240 CC./MIN.
n = 2.5 R.P.M.
θ = 35°
A
B
C

Fig. 3. Interpolation curves for Equation (23).

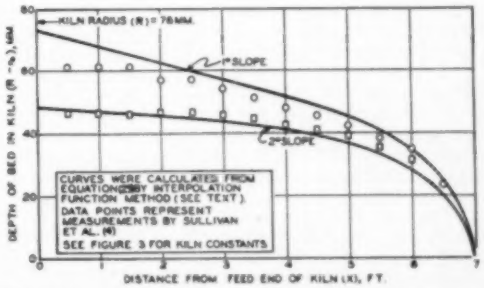
KILN RADIUS (R) = 76 MM.
1° SLOPE
2° SLOPE
CURVES WERE CALCULATED FROM EQUATION 23 BY INTERPOLATION FUNCTION METHOD (SEE TEXT).
DATA POINTS REPRESENT MEASUREMENTS BY SULLIVAN ET AL. (4)
SEE FIGURE 3 FOR KILN CONSTANTS
DEPTH OF BED IN KILN (R − r_o), MM
DISTANCE FROM FEED END OF KILN (x), FT.

Fig. 4. Comparison of calculated and measured bed depths for conditions that give significant droop in bed depth.

$$\frac{dr_o}{dx} = \frac{1}{77 - 12.5r_o} \quad (0 < r_o < 45 \text{ mm.}) \tag{26}$$

and

$$\frac{dr_o}{dx} = \frac{1}{2.5 + 2(7.5 - r_o)^2} \quad (45 \text{ mm.} < r_o < 75 \text{ mm.}) \tag{27}$$

That used for a 2° kiln slope was:

$$\frac{dr_o}{dx} = \frac{1}{1.6r_o + \frac{140}{r_o - 2.5} - 39} \quad (25 < r^o < 75 \text{ mm.}) \tag{28}$$

Equations (26)-(28) were then integrated to give expressions for depth of bed as a function of position in the kiln. Curves representing the integrated equations are shown in Figure 4, together with the observed values of the depth of bed reported in (*4*). Although Equation (23) is not valid at the discharge end, where the bed depth decreases rapidly, it is seen that the length of the section in which this rapid decrease occurs is so short that it is of little or no value to know either the length in which this occurs or the value of ψ over this length. For practical purposes, it suffices to set $r_o = R$ at the discharge end of the kiln. Inside the kiln, r_o approaches tangentially the value of r_o for uniform bed depth calculated by Equation (18) when $\psi = 0$. For practical purposes, the difference between the value of r_o obtained from Equation (23) and that from Equation (18) for $\psi = 0$ can be neglected as soon as the two values of r_o agree within practical limits, and it may be said that uniform loading is reached at that point in the kiln. The value of $R - r_o$ calculated for uniform loading for the 2° slope shown in Figure 4 is 52 mm., which is only slightly greater than the feed-end value, 48 mm., this indicates that uniform loading was almost reached. It was shown earlier that the maximum kiln capacity is reached with a depth of bed equal to the radius of the kiln. The capacity for uniform loading of the kiln used by Sullivan, et al. calculated by Equation (18) for a 1° slope is 142 cc./min. The rate used, 240 cc./min., was actually in excess of this, which means that ψ must take on a value of sufficient magnitude to give the required transport rate, as is shown by the plot of the theoretical equation in Figure 4. Since the depth of the bed is extremely sensitive to small fluctuations in experimental conditions when the depth of bed is nearly equal to the kiln radius, the differences between the observed and calculated values shown in Figure 4 are not considered excessively large.

Effect of Feed Rate on Time of Passage. Time of passage of material through a kiln is related to the feed rate and to the cross-sectional area of the bed by Equation (20)

$$t = \frac{LA}{q}$$

in which A and q are mutually dependent on r_o. Since dq/dr_o was shown by Equation (22) to diminish as r_o decreases, whereas dA/dr_o increases as r_o decreases, it is obvious that increases in the kiln feed rate, q, should result in increases in the time of passage and that these increases should be more pronounced at relatively high feed rates.

Only slight effects of kiln feed rate on time of passage were indicated by data (*4*). Comparison of the effects of feed rate indicated by these data with effects calculated for the same conditions by use of the theoretical equations is provided in Figure 5. This figure shows that there is reasonably close agreement between the observed values of time of passage at various feed rates and values calculated from Equation (23), which includes correction for droop in the bed. On the other hand, calculation of time of passage through the experimental kiln, assuming uniform loading, results in a curve (Fig. 5) that shows a relatively large effect of feed rate on time of passage. It is evident, therefore, that the conditions of the experiment by Sullivan, et al. were such that the effect of droop in bed depth largely counteracted the effect of feed rate on time of passage. It is expected that, with a relatively long kiln, the effect of droop on time of passage would be less and the relative effect of feed rate would be greater than the effects noted in the experimental kiln of the authors mentioned.

Time of Passage Through Kilns with Constrictions. For any rotary kiln operating under steady conditions, the volumetric transport rate, q, is the same at all points along the kiln length; thus, as is apparent from Equation (21), a constriction can alter the time of passage only by altering the bed cross-sectional area over a significant portion of the kiln length. Equation (23) is applicable to determination of the effects of several types of kiln constrictions on time of passage, as will be described below.

For thin, diaphragm-type constrictions and where the normal * depth of material at the point of the constriction would be less than the height of the constriction, the depth of bed immediately above the constriction is approximately equal to the height of the constriction, and the bed depth gradually decreases above the constriction. The value of r_o at the constriction is therefore equal to the radius of the constriction and serves as the terminal condition for solving Equation (23) for the depth of bed above the constriction until uniform loading is reached. The bed depth below a constriction of this type is not affected by the constriction, and time of passage through this section can be calculated as for a kiln without constriction. For thin constriction with a height less than normal bed depth at the point of constriction, it is obvious that there would be no significant effect on time of passage of material through the kiln since neither R nor r_o are altered over any appreciable distance.

A broad-type constriction alters the effective kiln radius for a relatively large portion of the kiln length; thus, the height of this type of constriction need not exceed the normal bed depth to have a significant effect on the time of passage. The time of passage through kilns equipped with such constrictions may be calculated by applying the theoretically derived equations separately to (a) the kiln section below the constriction, (b) the constricted section, and (c) the kiln section above the constriction. Calculation for the section below the constriction may be made as was described for kilns without constrictions. For the constricted section, Equation (21) may be used with R equal to the constriction radius and with the terminal value of r_o at the lower end of the constricted section equal to the smaller of either (a) the constriction radius or (b) r_o calculated for the uppermost point in the section below the constriction. Calculation of time of passage through the kiln section above the broad constriction may be made using R = kiln radius and using a terminal value of r_o at the lower end of the section equal to r_o calculated for the uppermost point in the constricted section. This terminal value of r_o exceeds the normal bed depth for this point in the kiln; thus, bed depth above the constriction will decrease to the value for uniform loading.

It is indicated from the analyses given above that changing the position of a kiln constriction should not affect time of passage through the kiln so long as the bed depth above the constriction is permitted to reach a constant value and so long as uniform loading exists below the constriction. If the constriction is placed so close to the feed end that uni-

* "Normal bed depth," as used here, refers to the bed depth that would exist without a constriction in the kiln.

form loading is not reached above the constriction, the lengthening of retention time by the constriction will be less than if it were placed further down the kiln. If the constriction is placed so close to the discharge end that it interferes with the normal droop, the lengthening of retention time by the constriction will be somewhat greater than if it were placed further up the kiln.

An example involving Equations (12) and (23), together with the general relations represented by Equations (20) and (21), deals with the calculation of the time of passage through an industrial nodulizing kiln in which the equivalent of a broad constriction exists. Sand and clay are heated to incipient fusion in the kiln to agglomerate the mixture into coarse aggregates or nodules. The kiln diameter is less in the combustion zone than in the remainder of the kiln due to the use of a thicker brick lining and the accumulation of partially fused material on the walls. The fused material is cut from the walls periodically with a heavy cutting bar. The dimensions and operating conditions of the kiln are:

Length, ft.	125
Shell diam., ft.	8
Inside diam., combustion zone, ft.	6.5
Thickness of fused layer (approximate), ft.	0.15
Inside diam., feed end, ft.	7
Length of 7-ft. diam. section, ft.	80
Rotation, rev./min.	3
Slope, in./ft.	0.5
Feed rate, cu.ft./min.	6.1
Angle of repose, degrees	45
Measured time of passage, min.	40

Since the conditions affecting the time of passage vary along the kiln, separate calculations must be made for each set of conditions affecting the time of passage. The time of passage for uniform loading through the 80-ft. section at the feed end can be calculated with Equation (12) since the bed is relatively shallow.

$$t = \frac{(80 \text{ ft.})(\sin 45)}{2\pi(3.5 \text{ ft.})(3/\text{min.})(1/24)} = 20.5 \text{ min.}$$

Equation (12) is also used to calculate the time through the 6.5-ft. diam. section in the combustion zone.

$$t = \frac{(45 \text{ ft.})(\sin 45)}{2\pi(3.1 \text{ ft.})(3/\text{min.})(1/24)} = .13 \text{ min.}$$

In the lower end of the 80-ft. section the bed builds up due to the reduced diameter of the combustion zone. The volume of this build-up divided by the volumetric feed rate gives the time which must be added to the two values calculated above to give the total time of passage. The extent of the build-up is calculated by Equation (23) using an interpolation formula for the solution, as exemplified by Equation (25). The depth of the bed in sections of the kiln where the bed depth is uniform is given implicitly by Equation (18), and is found to be 0.64 ft. for the 80-ft. section and 0.74 ft. for the 45-ft. section. The values of r_o for these two sections are 2.86 ft. and 2.36 ft., respectively. The variation of ψ upstream from a constriction varies only slightly in contrast to the strong variation associated with droop at the discharge end. Because of this, close approximations in the solution of Equation (23) are more easily made.

For this case it is adequate to determine ψ at three points. For values of r_o equal to 2.86, 2.66, and 2.36 feet the values of ψ (dr_o/dx) calculated from Equation (23) are 0, —0.017, and —0.031, respectively. An interpolation curve passing through these points and giving approximate values of dr_o/dx for values of r_o between 2.36 and 2.86 ft. is given by the formula

$$\frac{dr_o}{dx} = -\frac{1}{14 + \frac{9}{2.86 - r_o}}$$

The solution of this equation giving x as a function of r_o is

$$x = 119 - 14r_o + 9 \ ln \ (2.86 - r_o)$$

The build-up in the bed can now be plotted and determined graphically. It is found to be about 50 cu.ft. in excess of that for uniform loading. Since the volumetric feed rate is 6.1 cu.ft./min. the extension in time of passage due to this build-up is about 8 min. The total calculated time is therefore 41.5 min. and agrees closely with the measured value of 40 min.

Time of Passage Through Roto-Louvre Driers. Calculation of time of passage through Roto-Louvre type driers presents a special problem because of the tapered diameter characteristic of this type of drier. However, the theoretical relations that have been developed are readily adaptable to calculations relative to the Roto-Louvre-type drier as is illustrated in the following example.

Given: A Roto-Louvre drier with dimensions and operating conditions as follows:

Length, ft.	40
Feed end diameter, ft.	6.6
Discharge end diameter, ft.	10
Depth of bed at feed end, ft.	2.0
Feed rate, cu.ft./min.	12.5
Speed, rev./min.	3
Slope	Zero
Angle of repose of material, deg.	43

Sought: The relation of depth of material in the drier to distance from feed end of drier. (Such a relation would permit determination of average time of passage by application of Equation (21).)

The drier radius as a function of x is given by

$$R = 3.3 + 0.0425x \qquad (29)$$

Since r_o is expected to vary over a relatively small range only, it can be approximated as a function of x by a general power series of the type

$$r_o = A + Bx + Cx^2 + Dx^3 + \ . \ . \ . \qquad (30)$$

The derivative, dr_o/dx, is thus

$$\frac{dr_o}{dx} = B + 2Cx + 3Dx^2 + \ . \ . \ . \qquad (31)$$

If Equations (29), (30), and (31) are substituted in Equation (23), the values of the constants A, B, C, etc., can be evaluated at $x = 0$ by successive differentiation of the equation. This substitution gives (neglecting higher terms in the series):

$$(B + 2Cx + 3Dx^2 + \ . \ . \ .) = \frac{-\phi}{\cos\theta} + \frac{3q \sin\theta}{4\pi n \cos\theta}\left[\frac{1}{(3.3 + 0.0425x)^2 - (A + Bx + Cx^2 + Dx^3 + \ . \ . \ .)^2}\right]^{3/2} \qquad (32)$$

Since $\phi = 0$, the term involving it drops out. At $x = 0$ (feed end) $A = r_o$ or 1.3 feet. Substituting $A = 1.3$ in Equation (32) and solving for B, it is found that $B = 0.0331$. Differentiation of Equation (32) permits the calculation of

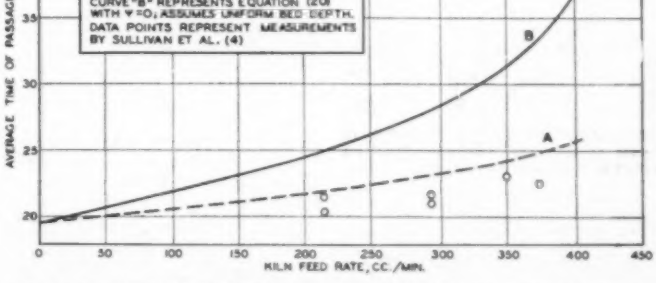

Fig. 5. Comparison of calculated and measured effects of kiln feed rate on time of passage.

C at $x = 0$ from the values of A and B already found. The value of D can be found likewise by differentiating a second time. Repetition of this procedure leads to a general series solution of the problem. For a converging series, as obtained in this case, the number of terms required depends on the accuracy desired. The solution found up to terms in x^3 is

$$r_o = 1.3 + 0.0331x - 0.000524x^2 + 0.00000663x^3 - \ldots \quad (33)$$

Since the value of the x^3 term at 40 ft. is only 0.4 ft. and since the series converges and is alternating, the error which results from neglecting terms higher than x^3 is less than 0.4 ft. in this case. The depth of bed at the discharge end is therefore

$$5 - (1.30 + 1.32 - 0.84 + 0.42 - \ldots) = 2.8 \text{ ft.}$$

This depth will, of course, exist only if the discharge end has a barrier to retain a 2.8-ft. bed in the drier. If the barrier at the discharge end is not this high, it will be the controlling radius in the drier and will serve as a starting point for calculating the depth in other parts of the drier. In actual operation a barrier 2.7 ft. high was needed at the discharge end to establish the conditions in the drier which were specified in the preceding problem.

For a second example involving the same drier with a feed rate of 9 cu.ft./min. and a barrier at the discharge end set to give a bed 2 ft. deep, the solution of Equation (32) will be repeated in a slightly different form which simplifies the work by reducing the number of terms required in the power series approximation. In the following calculation the series will be limited to three terms and the constants in Equation (32) will be represented by the letter F giving

$$\frac{dr_o}{dx} = B + 2Cx = F[(a+bx)^2 - (A + Bx + Cx^2)^2]^{-5/2} \quad (34)$$

The second derivative, d^2r_o/dx^2, is obtained by differentiating Equation (34) as follows:

$$\frac{d^2r_o}{dx^2} = 2C = -\frac{3F}{2}[(a+bx)^2 - (A + Bx + Cx^2)^2]^{-5/2} [2b(a+bx) - 2(B + 2Cx)(A + Bx + Cx^2)] \quad (35)$$

The constants B and C are obtained from Equations (34) and (35) by setting all factors of x equal to zero, giving

$$B = F[a^2 - A^2]^{-5/2} \quad (36)$$

and

$$C = -\frac{3F}{2}[a^2 - A^2]^{-5/2}[ab - AB] \quad (37)$$

Since the three-term series approximation is valid only for relatively small values of x, the error in the approximation must be estimated in order to insure calculations of sufficient accuracy. Since the series is found to be converging and alternating, the last term will indicate the approximate size of the error involved. The largest value of x used must therefore be limited, so that the last term does not exceed the tolerable error in the calculation. For values of x beyond the limiting value, a new set of constants may be calculated for the three-term series approximation. This step-wise procedure may be repeated as often as needed to extend the calculation over the full length of the drier. For the conditions of this problem the factors F, a, b, and A have the values 0.67, 5, 0.0425, and 3, respectively, giving values for B and C equal to 0.0104 and −0.0002, respectively. For the initial step the origin of x will be located at the discharge end and the distance towards the feed end will be considered negative. The initial series for r_o is,

$$r_o = 3 + 0.0104x - 0.0002x^2 \quad (38)$$

Values for r_o are 2.7 and 2.25 ft., respectively, for values of x equal to −20 and −40 ft. At −40 ft. (feed end) the x^2 term is 0.32 ft., indicating that the error in the approximation may be of this order. To obtain a more accurate value at the feed end the constants A, B, and C, can be recalculated at −20 ft. where the error in r_o is less than 0.08 ft. The new series with the origin for x located 20 ft. from the discharge end is

$$r_o = 2.7 + 0.0214x - 0.0004x^2 \quad (39)$$

For Equation (39) the value of r_o at x equal to −20 ft. (feed end) is 2.1 ft. and is in agreement with the measured value, which was 2.1 ft. also.

Summary

Theoretical equations for time of passage and bed depth in rotary kilns were derived from observations reported by Sullivan, et al. (4) together with observations made by the author on industrial kilns and Roto-Louvre driers. In the simplest case for lightly loaded kilns the time of passage is given by Equation (12),

$$t = \frac{L \sin \theta}{2\pi r n \phi}$$

and corresponds closely to the equation derived by the above mentioned authors on an empirical basis. For accurate calculations on heavy kiln loads the volume of the bed and time of passage must be calculated from the value of r_o given implicitly by Equation (18),

$$q = \frac{4\pi n}{3}\left(\frac{\phi + \psi \cos \theta}{\sin \theta}\right)(R^2 - r_o^2)^{3/2}$$

When ψ in Equation (18) is not equal to zero the problem involves the solution of the differential Equation (23),

$$\psi = \frac{dr_o}{dx} = \frac{3q \sin \theta}{4\pi n \cos \theta}(R^2 - r_o^2)^{-3/2} - \frac{\phi}{\cos \theta}$$

Several practical methods of solution for this equation are illustrated. The theoretical equations are corroborated by experimental data (4), together with data from industrial kilns and Roto-Louvre driers. The analysis clarifies previously obscure effects of the feed rate and constrictions on the time of passage through rotary kilns.

Acknowledgment

Acknowledgment is made to T. R. Mitchell, Jr., and Mildred Guernsey for making calculations and preparing drawings and to G. Hoffmeister for offering editorial assistance in the preparation of this paper.

Notation

(Applicable to any set of consistent units)

- A = cross-sectional area of bed
- c = length of chord at surface of bed
- D = kiln diameter
- L = length of kiln
- N = number of cascades per revolution
- n = speed of kiln
- q = volumetric transport rate in kiln
- R = kiln radius
- r = radius of particle path in bed
- r_o = minimum radius path in bed
- t = time, or time of passage through kiln
- u = average axial transport velocity of a single particle
- x = distance along kiln axis
- x_c = axial transport distance of particle per cascade
- $\theta°$ = angle of repose of material, degrees
- θ = angle of repose of material, radians
- $\phi°$ = slope of kiln, degrees
- ϕ = slope of kiln, radians
- ψ = angle between surface of material and kiln axis, radians

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Editor's Note: This paper was first submitted by Mr. Saeman in May, 1950, and a revised manuscript submitted in February, 1951. In mathematical derivations of this kind, matters of priority are often of importance.

PERFORMANCE OF SPINNING DISK ATOMIZERS

Part I

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An investigation was made of the performance characteristics of spinning disk atomizers. A comparison of the drop-size distribution from seven different types of disks operated at one feed rate and peripheral speed showed that disk design had a negligible influence on drop-size distribution for the conditions studied. Studies were made of the effect of disk speed, feed rate, and air pumping by the disk on power requirements and spray weight distribution. An equation was recommended for estimating power for disk atomizers.

A mathematical analysis of the velocity of liquid flow on spinning disks led to a nonlinear ordinary differential equation, which was solved for various conditions on IBM punched card machines. The solutions should be useful for predicting either liquid radial velocity on vaned disks or friction factors, depending on the type of experimental data available.

MANY chemical engineering processes involve the atomization of liquids. Typical of such processes are spray drying, spray absorption, spray crystallization, spray cooling, combustion, and spray reactions. The principal characteristic and advantage of atomizing is the creation of a large ratio of surface area to volume, resulting in increased heat- and/or mass-transfer rates, reaction rates, etc.

The atomizers found in industrial application usually may be classified as one of three basic types, namely, pressure nozzles, two-fluid nozzles, or spinning disks. The investigation reported in this paper was concerned with the characteristics of certain types of spinning disk atomizers which as a class have received little quantitative attention. Atomizers of this type have certain advantages over nozzles. For example, they have no small holes to clog; they can handle high rates of liquid feed; high pressures are not involved; and they are easy to clean. Moreover, they can usually be operated over a range of feed rates without requiring changes in other operating conditions. Among their disadvantages are high initial cost, high rotational speeds, and a spray pattern which covers a wide horizontal area. Rotating disk atomizers have found wide application in spray drying and spray cooling or crystallization processes.

Three important properties are usually associated with sprays and their formation. These are:

1. *Drop-size Range and Uniformity, and the Factors Influencing Them.* Information on this property is important in the control and operation of equipment. For example, if a wide drop-size range exists in a spray dryer, small drops may dry rapidly and overheat before the larger drops are dry. Further, overall drying time—and hence equipment size—depends on the size of the largest drops present. Finally, the physical properties of a spray dried product, such as bulk density, color, solubility, dustiness, etc., are influenced by particle size and distribution.
2. *Manner in Which a Spray Is Distributed.* This property is usually designated as weight distribution and is concerned with the direction and the distance drops are thrown by an atomizer, and the amount or weight of spray falling at various points in the spray area. Data on this property are essential to establishing the size and shape of spray process equipment.
3. *Power Required for Atomization.* Data on power requirements are essential to the proper selection of motor or turbines for atomizing at specified rates of flow.

This paper is a report of a study to determine and compare the drop-size distributions of sprays produced by seven different types of rotating disk atomizers for one feed rate (25 lb./min.) and one peripheral speed of the disks (12,500 ft./min.) in order to determine whether any of the disk designs studied showed an unusual or marked atomizing ability compared with the others. Other objectives were to obtain power and weight distribution correlations for various operating conditions. A theoretical phase of the research consisted of a mathematical analysis of the flow characteristics of the liquid on a spinning disk.

Previous Work

Disk Design. The earliest type of atomizing disk was a smooth, flat plate. But, since this type of disk permitted excessive slippage of the liquid over its surface, smooth disks with curved surfaces were designed to increase the friction between liquid and disk and reduce slippage (*25, 36, 55, 37*). Projections, elevations, depressions, fillets, ribs, grooves were next added to improve atomization (*11*). These designs, however, have been said to produce nonuniform sprays (*3, 25, 36*). Double and triple bowl rotors were next developed (*38-42*). Impeller blades or vanes have been used to produce fine atomization (*31*). In this design, the vanes are perpendicular to the disk surface and to a cover plate, and the liquid travels along the vane in a thin film perpendicular to the disk and cover plate. (Fig. 1.) Vanes have been made in various shapes and sizes, from straight, radial vanes to those with double curvature.

It has been claimed that curved vanes will give a more uniform and continuous film, which should produce a uniform drop size (*31*). Forward leading vanes have been used to increase the liquid's peripheral speed (*56*), although this would certainly be done at the expense of radial velocity.

Drop-size Distribution. The literature contains few data on drop-size distribution for rotating disk atomizers. Hartmann (*15*), in an unpublished thesis, obtained particle-size distribution for spray-cooled soaps from a disk with radially drilled holes for liquid flow.

Only qualitative statements may be found in the literature on the effects of disk design, disk speed, feed rates, and liquid

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† Present address: Eastman Kodak Co., Rochester, New York.

Part II of this paper will be run in an early issue of Chemical Engineering Progress.

Table 1 in its entirety is on file (Document 3365) with American Documentation Institute, 1719 N Street, N.W., Washington 6, D. C. Obtainable by remitting $1.00 for microfilm and $1.00 for photocopies.

properties on the degree of atomization. Thus, particle size has been said to vary inversely with disk peripheral velocity (*46*). Increased feed rate according to Hartmann (*15*) increases drop size. At low rotational speeds and feed rates, Walton and Prewett (*54*) found drop size to be substantially independent of feed rate on smooth, saucer-shaped disks. These workers obtained drop-size distribution data for small disks operating at very low liquid feed rates; their results indicated a high degree of drop-size uniformity under these conditions.

Wallman and Blyth (*53*) investigated the characteristics of spray-dried sodium silicate produced by a disk atomizer under various conditions of dryer operation. Using smooth saucer-shaped and vaned disks, they studied the effect on particle-size distribution of feed temperature, air temperature and velocity, liquid surface tension and viscosity, and disk speed. Particle-size distribution was determined by a combination of sieving and sedimentation.

Mechanism of Atomization. The mechanism of atomization has been explained as the extension of a liquid surface to an unstable configuration which collapses to form drops. This unstable configuration may be in the form of films or ligaments. A theory of atomization by air friction and liquid turbulence has been postulated by many workers (*3, 14, 22, 44-47, 49*) from studies of atomization with nozzles. Castleman (*6-9*) proposed a theory of atomization by the collapse of ligaments, and considered ligament formation as an essential step in the process.

Bär (*2*) distinguished between two mechanisms of atomization, namely, atomization by surface forces or air friction, and atomization by inertia or mass force.

Hinze and Milborn (*18*) postulated three types of mechanisms: (1) disintegration by direct drop formation, (2) disintegration by ligament formation, and (3) disintegration by film formation. These mechanisms are discussed further under Theoretical Considerations.

Drop Sampling Techniques. The average drop diameter of a spray can sometimes be determined by a photometer method (*44-46*) or by a corona method, methods which do not involve taking physical samples of the drops. In the former method the percentage absorption of light by a spray is used to calculate a mean droplet diameter. Refraction, reflection, and shading of some drops by others tend to make this method inaccurate (*50*). In the corona method, use is made of the fact that a point source of light viewed through a uniform mist will show diffraction rings. From the ring order, angle, and the wave length of the light used, the droplet diameter can be calculated. However, for sprays with a distribution of drop sizes, the usual case, the corona method gives an inconclusive result (*20, 22*).

Photographing a spray in operation with a thin sheet of light for illumination has been suggested (*20*). However, focusing problems and disturbances from the unlighted drops in the spray have usually made this approach impractical.

The usual method for obtaining drop-size distribution data is to collect a sample of the spray and in some manner measure and count the particles. The most convenient method would be the spraying of a liquid that could be collected as a solid and analyzed by conventional methods. This could be accomplished by actually drying a solution (*15, 53*) or by using a molten material which would solidify rapidly after

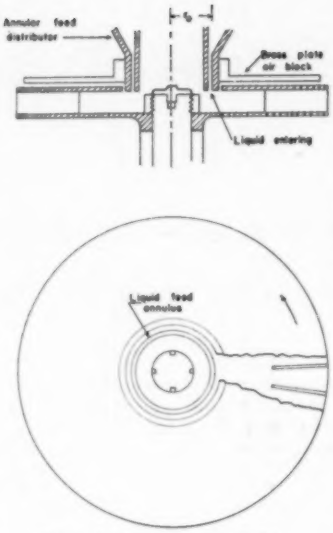

Fig. 1. Liquid flow on a vaned disk.

atomizing. In this connection one method reported consisted of freezing the droplets by collection on a liquid cooled to below the freezing point of the droplets (*27*). With these techniques, sieving or sedimentation is required to separate the size fractions.

Several methods have been reported for collecting drops on glass slides coated with grease, oil, soot, or magnesium oxide (*19-21, 26, 30*). The samples are either photographed and counts made from enlarged pictures, or the drops on the glass slide are counted directly under a microscope.

When a glass slide is coated with soot or magnesium oxide, the droplets penetrate the surface of the coating, leaving holes which are related to the drop sizes (*18, 22, 24, 52*). Most users of this technique have assumed the hole diameter to be equal to the drop diameter. Stoker (*51*) showed that in the case of soot-coated slides the difference between the recorded and true diameters was a function of the Weber number. He presented data which checked his theoretical equation.

Dyed drops have been collected on blotting paper and the resulting stain correlated with the drop size (*52*). This is suited primarily to large drops in a sparse spray.

The collection of the spray sample in a so-called "immersion-cell" containing an immiscible solvent has been used (*10*). Schmidt (*48*) suggested using castor oil as the solvent, although with this fluid the drops hung from the upper surface, and small drops had difficulty penetrating the surface of such a relatively viscous oil (*43*). Rupe (*43*) investigated the immersion-cell technique in considerable detail.

Power Requirements and Liquid Velocity. The power requirements of a rotating disk atomizer depend on the speed of rotation, liquid feed rate, and liquid properties (*36*). Phillip (*33*) presented some plots of gross power vs. feed rate for three rotary speeds on a Kestner disk. Unfortunately, the power used to overcome friction and windage was so appreciable in his setup that an accurate determination of net power from the curves was impossible. Further, the design of the disk was not reported.

Marshall and Seltzer (*29*) gave an idealized relationship for power consumption by spray disks, assuming no friction or liquid slippage and that liquid was fed at the center of the disk.

Only slight consideration has been given to the velocity of the liquid over the disk surface. Hartmann (*15*) assumed friction to be negligible i.e., that theoretical velocities were attained. Bär (*2*) developed equations for liquid flow on smooth disks and a type of vaned disk. Hinze and Milborn (*18*) developed similar equations for a smooth disk. Marshall and Seltzer proposed more rigorous equations to account for slippage and friction. Complete solutions were not included, however.

Theoretical Considerations

Spray-Weight Distribution. A rotating disk distributes its spray horizontally in the general shape of an open umbrella. The small droplets soon lose most of their horizontal velocity, and gain a vertical velocity component under the influence of gravity. Under ideal conditions, the work of Lapple and Shepherd (*23*) would be useful in calculating theoretical particle trajectories. Unfortunately, in actual operation air currents and turbulent mixing preclude application of their theoretical predictions. Three factors cause air circulation: (1) the air-pumping action of the disk, (2) frictional drag of the outer disk surfaces on the air, and (3) the transfer of momentum to the air by the sprayed liquid. These complicating factors make it difficult to postulate a quantitative expression to predict the weight distribution for disk atomizers.

Drop-size Distribution and Mechanism of Atomization. It is difficult to derive a useful mathematical expression to predict drop-size distribution from the magnitudes of the various operating variables involved in disk atomization. It is only possible, at best, to predict qualitatively how drop size will vary with changes in disk speed, disk diameter, disk design, feed rate, and liquid viscosity, density and surface tension. Consequently, at this time theoretical treatment of drop-size distribution must be restricted to a general consideration of the factors influencing the formation of drops.

Bär (*2*) distinguished between two ways of producing a spray with disks: (1) atomization by surface forces at a liquid-air interface, and (2) atomization by mass forces, or the tearing apart of the liquid by centrifugal or gravity forces. The former was termed a velocity spray, the centrifugal force serving only to give the liquid velocity, with atomization occurring by friction with the air. In this type of spray, it was

claimed that drop size would be inversely proportional to the square of the peripheral velocity. When mass forces predominated, Bär claimed that air friction would have no effect and that drop size would be more uniform and inversely proportional to the square root of the centrifugal force, or to the first power of the peripheral velocity for any one disk. Wallman and Blyth (*53*) reported drop size to be inversely proportional to centrifugal force. Bär developed equations for calculating the diameter of the largest drops for both types of atomization and compared them with experimental data to obtain an empirical coefficient for the equation for velocity spraying. Bär also observed that both types of atomization could occur simultaneously. He collected large drops in a narrow size range with some accompanying small satellite drops. Similar results were obtained by Walton and Prewett (*54*) for low feed rates.

Hinze and Milborn (*18*), using small cup-shaped disks operated at low speeds and low feed rates, obtained excellent photographic evidence to support their proposed three mechanisms: direct drop formation, ligament formation, and film formation. In atomization by direct drop formation, the equivalent of Bär's disintegration by mass forces, the drops took shape in bulges on the disk periphery, and were thrown off singly. With increased feed rates these bulges formed thin jets which became unstable and eventually broke up at some distance from the disk. When still larger feed rates were used, the jets joined to form a continuous film extending a distance out from the disk edge before breaking up irregularly into ligaments and globs of liquid. These last two mechanisms would be equivalent to Bär's velocity spraying. Uniformity of drop size was said to decrease as the mechanism changed in this manner. It was recommended that in order to obtain uniform films and ligaments, and hence uniform drop size, the following conditions should be met: (1) the centrifugal force should be large compared with the gravitational force, (2) disk rotation should be vibrationless, (3) liquid feed rates should be uniform, and (4) disk surfaces should be smooth.

Hinze (*18*) and Walton and Prewett (*54*) used the same general operating conditions and found high uniformity of drop size at low, uniform feed rates when the disk was completely wetted and no ligaments or films were formed. When ligaments were present, they reported two principal particle-size ranges. The lower range probably consisted of the small, satellite droplets created during the collapse of ligaments as reported in the pioneer work of Rayliegh (*35*).

To summarize, the drop-size distribution from spinning disk atomizers apparently depends on rotational speed or centrifugal force, depending on the magnitude of the feed rate. At low feed rates and low disk speeds, mass forces predominate, and the drop size is primarily an inverse function of centrifugal force. At high feed rates and/or high disk speeds, the drop size is primarily an inverse function of peripheral speed. Undoubtedly, an intermediate range of feed rates and disk speeds exist wherein drop size is influenced equally by both centrifugal force and peripheral speed. The liquid properties —such as density, viscosity, and surface tension—probably serve, primarily, to influence the range in which centrifugal force and disk speed predominate.

Power Requirements. The theoretical net power for disk atomizers based on the kinetic energy to give the liquid a certain resultant velocity as it leaves the disk can be calculated from the relation

$$P = \frac{wv^2}{2g_c} \qquad (1)$$

where

P = net power
w = liquid feed rate
v = resultant liquid velocity as it leaves the disk
g_c = factor to convert pounds mass to pounds force

In order to develop this equation in terms of disk speed, further assumptions must be made regarding the liquid velocity.

When no circumferential slippage occurs, the liquid may be assumed to attain the peripheral velocity of the disk. The liquid radial velocity, however, depends upon its radial position, the opposing frictional drag at the disk surface, the rate of rotation, and the point of liquid deposit on the disk.

If the liquid is deposited at the center of the disk and friction is negligible, it can be shown that the two components of velocity are equal. Since these components are perpendicular the resultant velocity is given by the following equation:

$$v = \sqrt{2\omega r} \qquad (2)$$

where

r = disk radius
ω = angular velocity of disk

Substituting in Equation (1), the power in kilowatts is given by

$$P_k = 7.70(10^{-9})w(Nr)^2 \qquad (3)$$

where

P_k = net power required, kw.
N = rotational speed, rev./min.

When the liquid is fed at $r = r_e$, Equation (3) becomes

$$P_k = 7.70(10^{-9})wN^2[r^2 - (\tfrac{1}{2})r_e^2] \qquad (4)$$

The exact equations for velocity allowing for friction are too complicated to substitute conveniently into the power formula.

Liquid Flow on Disk Atomizers. Liquid flowing along a rotating disk is opposed by frictional drag. Frequently, friction is ignored as being negligible (*4*). Hartmann (*15*) did this, basing his decision on observations of the path

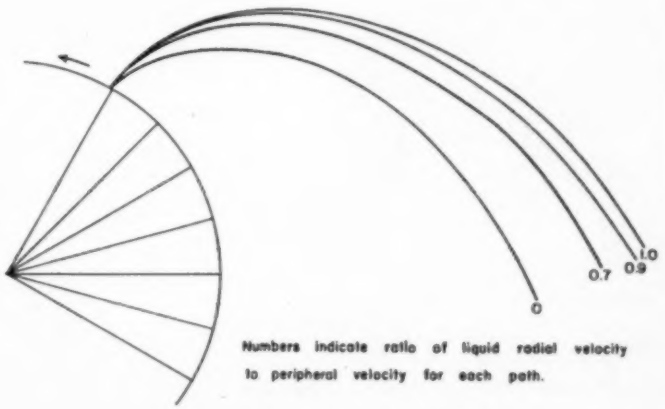

Fig. 2. Flow pattern of liquid leaving a vaned disk.

the liquid took upon leaving the disk. His photos showed a great similarity to the idealized flow patterns; but they would have the same appearance even with appreciable friction, since quantitative changes in the pattern are very difficult to detect, as shown by the curves in Figure 2.

Bär (2) developed an equation to include frictional drag, and assumed radial velocity equal to the theoretical value, ωr, to permit integration. Thus, his result was only a first approximation. In developing an equation for smooth disks, he assumed that the frictional retarding force was in equilibrium with centrifugal force. This equation indicated that the radial velocity would be inversely proportional to the cube root of the radius, an improbable conclusion. The assumption of balanced forces was not valid for all radial positions, but might apply as a limiting condition, as will be demonstrated later.

The equations for liquid velocity on rotating disks, of the form originally presented by Marshall and Seltzer (29), were solved in this study for various values of the parameters.

For viscous flow on a radial-vaned disk, a differential equation for the radial velocity of the liquid can be developed from a force balance on an element of fluid (Fig. 3) as follows:

Centrifugal force—frictional drag = mass × acceleration. Symbolically this may be written:

$$(mb\ dr)\rho\omega^2 r - \mu b\ dr \left(\frac{dv_r}{dy}\right)_{y=0} = (mb\ dr)\rho \frac{d(v_r)_{ave}}{dt} \qquad (5)$$

For a parabolic velocity gradient through the liquid film:

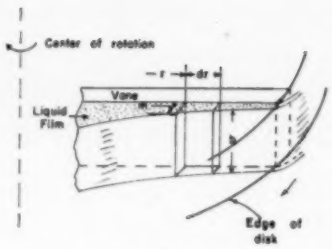

Fig. 3. Liquid flow on a vane.

$$\left(\frac{dv_r}{dy}\right)_{y=0} = \frac{3(v_r)_{ave}}{m} \qquad (6)$$

where $(v_r)_{ave}$ = the average radial velocity of the liquid at r. For convenience, the subscript *ave* will be dropped in the remaining development, since point liquid velocities at various depths in the film are no longer involved.

Introducing the continuity relation, $m = q/bv_r$, and substituting Equation (6) into Equation (5), the following nonlinear ordinary differential equation for laminar flow on a vaned disk is obtained:

$$v_r\left(\frac{dv_r}{dr}\right) + \left(\frac{3\mu b^2}{\rho q^2}\right) v_r^3 - \omega r^2 = 0 \qquad (7)$$

The area of liquid contact along the upper and lower disk surfaces was assumed negligible compared to that along the vane, that is, b was assumed equal to $(b + 2m)$.

The corresponding equation for turbulent flow is

$$v_r\left(\frac{dv_r}{dr}\right) + \left(\frac{fb}{2q}\right) v_r^3 - \omega r^2 = 0 \qquad (8)$$

where f = the usual friction factor for turbulent flow.

It can be seen that Equations (7) and (8) are of the general form:

$$v_r\left(\frac{dv_r}{dr}\right) + Av_r^3 - \omega r^2 = 0 \qquad (9)$$

in which the coefficient A depends on the type of flow and the nature of the velocity gradient across the liquid film. It is not always possible to establish the correct value of A, but a range of values can be estimated from the range of operating variables and checked with an experimental value.

Equation (9) was solved, for various values of A, corresponding to Equation (7), and r_{ov} on the *IBM 602A* calculator by the University of Wisconsin Computing Service. The results are presented in a later section.

By a similar development, the equation for flow on a smooth disk without slippage is

$$v_r\left(\frac{dv_r}{dr}\right) + A'rv_r^3 - \omega r^2 = 0 \qquad (10)$$

where

$$A' = \frac{12\pi^2\mu}{q^2\rho}$$

Experimental Equipment and Procedure

Experimental atomizing disks mounted on a vertical shaft held in a C-frame were belt-driven by a 1 hp., 3600 rev./min. motor. A speed range from about 450 to 10,500 rev./min. was obtained with various pulley combinations. The whole assembly was mounted on an angle-iron frame, such that the disks were about 4½ ft. above the floor. Disk rotation rates were determined by a Strobotac.

Water was pumped to the disk through a rotameter. A by-pass line from the gear pump outlet back to the tank permitted accurate adjustment of the feed rate over a range of 10-70 lb./min. (Fig. 4).

Liquid was introduced through an annular feeder (Fig. 1) to provide a uniform distribution of the feed on the disk. A large reservoir above the annulus ensured uniform distribution at low feed rates.

Eighteen different disks were tested. Eleven of these were variations of the radial-vaned type, the rest were of the curved vane, homogenizing, and bowl and saucer varieties. Photographs and sketches of the different types are shown in Figures 5 to 12.

Determination of Weight Distribution. The spray-weight distribution was determined by means of a collecting pan in the shape of an 18° sector of a 10-ft. radius circle. The pan was divided into ten compartments 2 in. deep at 1 ft. radial intervals. It was positioned below the disk with the center of the sector circle directly under the center of the disk. Drains from each compartment emptied into graduates.

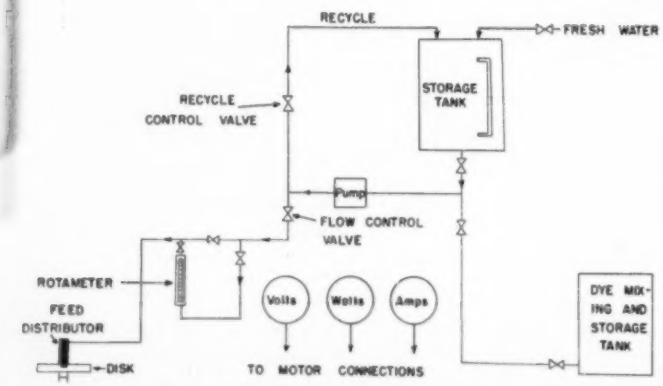

Fig. 4. Liquid feeding system for rotating disk atomizers.

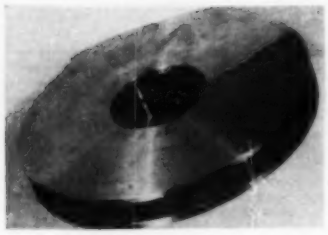

Fig. 5. Radial vaned disk.

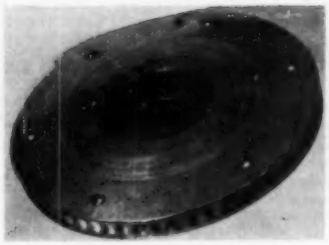

Fig. 6. Disk B-1.

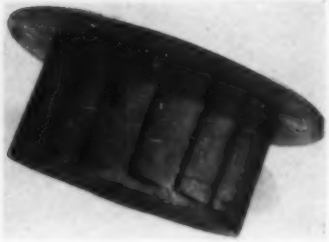

Fig. 7. Disk B-2.

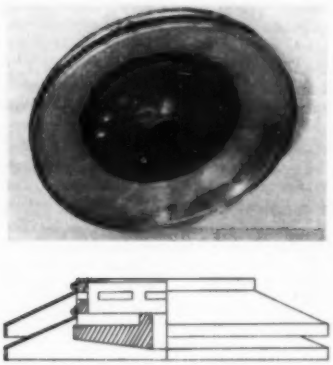

Fig. 8. Disk B-3.

Since the pan was exactly 1/20 of a circle and of large radius, it should have collected 1/20 of the total liquid sprayed during the run. This was checked from the feed rate and run duration, and agreement was always within 3%. Considering that no liquid was collected in the zero to one ft. interval, due to a shield over the motor, and that some of the spray could go farther than 10 ft. from the disk (both amounts proved to be negligible), this was considered to be satisfactory agreement. It also indicated uniform distribution of the feed to the disk, negligible loss of liquid by evaporation, and negligible effect of walls and fixtures in the room. After allowing a 5 min. drainage time, the quantity of liquid collected in each compartment was recorded.

The data could be reproduced with a maximum variation of about 3%. Usually two runs were made for each spray and the results averaged. More than 200 runs were made with different disks, feed rates, and disk speeds to determine the effect of these operating variables on the weight distribution pattern. The data were also used to determine the positions for taking drop samples, and for weighting drop-size distribution data.

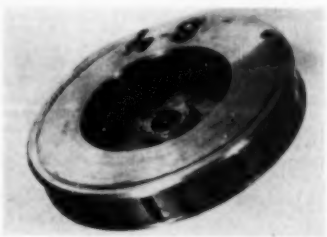

Fig. 9. Disk B-4.

Determination of Drop-size Distribution. Drop-size distributions were determined for the sprays from the seven disks of different design shown in Figures 5 to 11. The disks were operated at a single feed rate and approximately equal peripheral velocities. The test conditions are listed in Table 1.

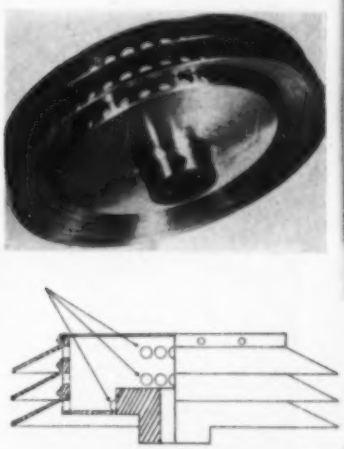

Fig. 10. Disk B-5.
Holes around whole circumference.

Fig. 11. Disk B-6.

The immersion-cell method of drop sampling used by Rupe (*43*) was adopted in order to take high contrast photographs of the drops when they were in a spherical shape. The sampling cells used were 1 in. inside diameter, and ¼ in. deep (Fig. 13). In order to prevent flattening of the drops, it was necessary to coat the cell bottom, made of optically flat glass, with GE Dri-film 9987, a nonwetting agent. The immersion fluid was Stoddard solvent. Rupe's work indicated that this method of collection gave spherical water droplets resting on the bottom of the cell, which could be easily photographed by transmitted light.

It has been shown (*20*) that when sampling particles in a moving air stream drops smaller than 20μ are frequently lost and, in fact, no drops smaller than 5μ can be collected unless very small sampling slides are used, or very high fluid velocities are employed. This difficulty was not present in this study, since sampling was done in a region of essentially zero air velocity around the sampling cell so that the drops fell into the cell at their terminal settling velocities.

When drop samples were taken, an aqueous solution of 3% nigrosine dye was sprayed. This amount of dye ensured opaque photographic images of drops with no highlights whatever. This was desirable for ease in discerning small droplets and was requisite to the development of an automatic counter. The aqueous dye solution had the same properties as water, except for a slightly lower surface tension. A comparison of the weight distribution at a given set of operating conditions for water and the nigrosine solution showed differences only within the range of experimental error.

The drop sampling procedure was as follows: The immersion cells were cleaned by repeated flushings with Stoddard solvent and inspected for cleanliness by observation of their image on the ground glass screen of a photomicrographic camera, which was used also for photographing the samples. When the cells were cleaned they were filled with the immersion fluid and covered with caps numbered for cell identification.

The sample cells were placed 36 in. below the plane of the disk, the same distance at which weight distribution data were taken. Seven sample points were chosen, extending from 2 to 9 ft. from the disk, so as to represent the weight distribution pattern completely, as shown in Figure 14. Two cells were placed at the positions where the spray was densest. Thus, ten cells were used for the seven sample points.

The cells were supported on a thin rail so as to disturb the spray as little as possible. After the spray was allowed to run for a short time to ensure air saturation, the experimenter carefully exposed each cell to the spray. This operation required extreme care to avoid disturbance of the general flow of the spray, and to ensure the correct exposure time, since this varied with the position of the sample. Judgment of exposure time became a matter of experience. Overexposure of a cell resulted in coalescence of the drops. Underexposure resulted in too few drops for the required accuracy. The cells were positioned so that, if necessary, one cell could be omitted without destroying the overall results.

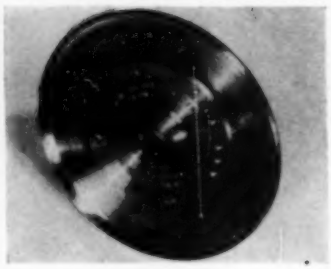

Fig. 12. Disk B-7.

The droplets in the cell were magnified and photographed with a photomicrographic camera. Before photographing, the cells were completely filled with solvent and also covered with a piece of optically flat glass to eliminate the distortion caused by a meniscus. It was noted that standing for a 12-hr. period made no difference in the drop-size distribution in a cell. As many as 14 different pictures were taken of the drops in one cell.

A photographic film was used, which gave good contrast, with a completely black background as shown in Figure 15.

A duplicate sampling run was made for one spray, and indicated satisfactory reproducibility. The drop-size-distribution curves closely duplicated each other and the average drop diameters agreed to within 1% on volume and surface area bases.

Errors may have occurred in the photographing process, in setting of magnification, either in the calibration of the stage micrometer or in the measuring of the enlarged image. By photographing a stage micrometer at the desired magnification and measuring the developed image, it was proved that the magnification recorded on the developed negative was the same as that measured on the focusing screen. Errors in the drop-size distribution resulted from the fact that only a finite number of drops were counted per sample and that the size classes in which they were classified had a finite width.

For an average of 6,000 drops counted for each spray, it could be expected that with a probability of 0.95 the absolute error in a cumulative numerical representation would be less than 2%. This error was decreased by sampling at selected and weighted positions. Any drops not present in large numbers, such as largest and smallest, were not as well represented as those present in large numbers and hence would be more uncertain.

TABLE 1

Operating Conditions for Drop Sampling

Disk type	rev./min.	Disk diam., in.	Periph. vel. ft./min.	Disk loading
B-1	6700	7	12,280	Vanes full
B-2	9730	5	12,720	40% full
B-3	6860	7	12,570	2 bowls, 40-60%
B-4	9950	5	13,010	50% full
B-5	6730	7	12,320	2 bowls, 50-50
B-6	7800	6	12,320	25% full
V-2	9750	5	12,750	Vanes full

Liquid sprayed: 3% nigrosine solution in water.
Surface tension of this solution was 63 dynes/cm.
Other properties were the same as those of water at 16°C.

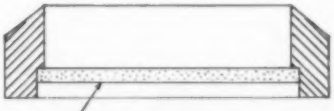

Fig. 13. Spray sampling cell.

Small errors in measuring drops and in recording counts undoubtedly occurred. However, it was felt that these errors were always less than the error caused by the limited size of the sample collected.

Power Measurement. Gross power input was measured with a wattmeter. Windage, belt and bearing friction, and copper and core losses in the motor were deducted from the measured power to give the net power delivered to the liquid. Power for windage and friction requirements for a given setup was measured at no load and assumed constant when feed was introduced. The copper losses in the motor were determined from the current measured for the test, and a curve of copper losses vs. current obtained from a locked rotor test on the motor. The net power was then calculated from the wattmeter reading by subtracting the two correction terms.

Power measurements were made for different disk designs, feed rates, and disk speeds. In addition, effects of pumping air, of linear velocity of the feed, and of direction of the disk rotation (for curved vane disks) on power consumption, were observed.

Calculation and Correlation Procedures

Weight Distribution. Weight-distribution data consisted of a record of the volume of liquid collected in a given time in each compartment of the sector-shaped pan.

The amount collected in each compartment was converted to a fraction of the total spray by dividing it by the total amount collected. These fractions were plotted as first approximations, as per cent per foot vs. the radial distance to the midpoint of the compartment; i.e., a differential distribution plot.

The total percentages collected up to various radial distances were next plotted against distance to obtain an integral or cumulative curve. The points on this curve involved no assumption as to position, as in the case of the differential curve. A graphical differentiation of the cumulative plot thus gave more precise point values for the differential curve. A comparison of the differentials of the cumulative plots with the plot of the original data, using the midpoints of the pan compartments, showed good agreement, well within the experimental error. Whenever weight distribution data were used, they were taken from the cumulative curves.

Procedure for Calculating Drop-size Distribution from Experimental Data. From the photographs of the spray samples for a given spray, the number of drops in each of 20 equal size classes for each of the seven sample points was obtained by measuring, counting, and classifying the drop images.

Since there was usually a smaller number of drops in the large size classes, and since the large drops contributed appreciably to the distribution on a volume basis, the curves were smoothed for these drop sizes and the adjusted values used. This could be done without loss in accuracy.

When the numerical fractions of drops had been adjusted for each of the seven sample points, the next task was to combine the seven point distributions into one over-all distribution for the spray.

It was possible to combine individual sample point distribution data into one over-all distribution on the basis of number, surface area, or volume with equivalent results. All drop-size distribution data presented in this paper were combined while in the numerical distribution form. Therefore, each sample point was weighted on the basis of the numerical fraction of drops it represented. The set of weighting factors required was calculated from the weight distribution of the spray and the drop-

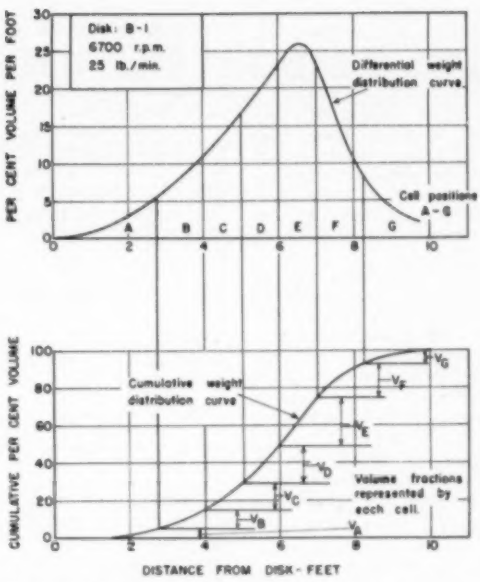

Fig. 14. Positions of drop sampling points and method of weighting.

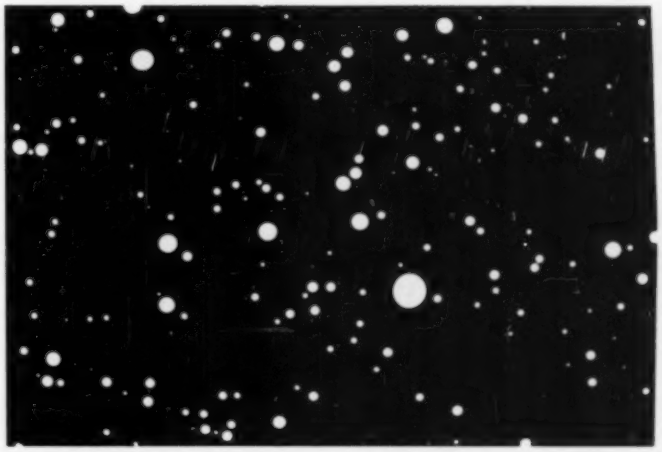

Fig. 15. Specimen photograph of drops in sampling cell taken with transmitted light.

size distribution data for each sample point. From the latter, the average volume of a drop at each point was calculated. Dividing these into volume fractions represented by the corresponding sample points (Fig. 14) yielded the average number of drops at each position per unit volume of spray. The fraction of the sum represented by each of these was the numerical weighting factor. With these factors the seven sample points were combined to give the numerical drop-size distribution for the entire spray.

Drop-size-distribution data may be based on number, volume or surface area. Each basis is useful for a different purpose. Volume and surface area distributions can be calculated from the numerical distributions by use of the average volumes and surface areas of each size class and by assuming uniform distribution in each class.

Average drop diameters also can be calculated on any of several different bases. The numerical average diameter defined by

$$d_n = \frac{\sum^{k}(n_i d_i)}{\sum^{k} n_i} \qquad (11)$$

can be obtained directly from the numercial drop-size distribution data as follows:

$$d_n = F_1 D_1 + F_2 D_2 + F_3 D_3 + \ldots F_i D_i + \ldots = \sum^{k} F_i D_i \qquad (12)$$

where F_i is the numerical fraction of drops in the ith size class, and D_i is the mean diameter of the ith size class. The average diameter, d_s, based on surface area is given by

$$d_s = \sqrt{\frac{\sum^{k}(n_i d_i^2)}{\sum^{k} n_i}} \qquad (13)$$

where n_i = number of drops of diameter d_i. In this study d_s was calculated from the following expression:

$$d_s = \left[\frac{F_1 s_1 + F_2 s_2 + F_3 s_3 + \ldots F_i s_i \ldots}{\pi}\right]^{1/2} = \sqrt{\frac{\sum^{k}(F_i s_i)}{\pi}} \qquad (14)$$

where s_i is the average surface area of drops in the ith size class.

The average diameter, d_v, based on volume is given by:

$$d_v = \sqrt{\frac{\sum^{k}(n_i d_i^3)}{\sum^{k} n_i}} \qquad (15)$$

In this study, d_v, were determined as follows:

$$d_v = \left[\frac{6}{\pi}(F_1 v_1 + F_2 v_2 + \ldots F_i v_i \ldots)\right]^{1/3} = \sqrt[3]{\frac{6\sum^{k}(F_i v_i)}{\pi}} \qquad (16)$$

where v_i is the average volume of drops in the ith size class.

An average diameter based on the ratio of volume to surface area can also be calculated. This would be the diameter of the drop whose volume to area ratio would be the same as that of the entire spray, expressed by

$$d_a = \frac{\sum^{k}(n_i d_i^3)}{\sum^{k}(n_i d_i^2)} \qquad (17)$$

or, as was used in this study,

$$d_a = \frac{6\sum^{k}(F_i v_i)}{\sum^{k}(F_i s_i)} = \frac{d_v^3}{d_s^2} \qquad (18)$$

Three types of mean diameters, D_n, D_s, and D_v, can be read directly from the 50% point of the cumulative drop-size plots based on number, surface area, and volume respectively. The relationships between these average and mean diameters have been discussed by various writers (*1, 12, 13, 16, 17, 28, 32, 34, 57*).

Power and Liquid Velocity Calculations. The net power utilized in atomizing the liquid was calculated by subtracting from the measured input power, the friction and windage loss for the disk, and the copper and core losses for the motor. Thus,

$$P_k = P_{in} - P_{wfc} - P_{cu} \qquad (19)$$

where

P_k = net power utilized
P_{in} = gross power, measured
P_{wfc} = power consumed in windage, friction, and core losses
P_{cu} = power consumed in copper losses

The velocity of the liquid leaving the disk could be calculated from the net power only if all this power went into the kinetic energy of the liquid. (Energy for the creation of new surface was estimated to be less than the experimental error.)

Other Calculations. The thickness of the liquid film at any point on the disk can be calculated by application of the continuity equation. For smooth or vaned disks:

$$q = v_r A \qquad (20)$$

where

q = volumetric feed rate
A = film cross-sectional area
v_r = radial liquid velocity

For vaned disks in particular

$$m = \frac{w}{\rho c v_r b} \qquad (21)$$

where

m = film thickness at r
w = weight rate of feed
ρ = liquid density
c = number of vanes
b = vane height

The efficiency of an atomization process defined as the fraction of the net power used to create new surface, can be calculated from the expression:

$$E = \left(\frac{\sigma A}{P_k}\right) 100 \qquad (22)$$

where

E = efficiency
σ = liquid surface tension, dynes per cm.
A = surface area created per unit time
P_k = net power imparted to liquid

or

$$E = 0.283\,\frac{\sigma w}{P_k d_a \rho} \qquad (23)$$

The rate of surface creation may be obtained as follows:

$$R = 1.83(10^8)\,\frac{w}{d_a \rho} \qquad (24)$$

where

R = rate of surface creation, sq.ft./min.
w = liquid feed rate, lb./min.
d_a = average drop diameter on a volume to surface basis, microns
ρ = liquid density, lb./cu.ft.

(Acknowledgment, Notation, Literature Cited, and Discussion will run in Part II.)

MASS-TRANSFER RATES ON BUBBLE PLATES

FUNDAMENTAL FACTORS

PART I

J. A. GERSTER, W. E. BONNET †, and IRWIN HESS ‡

University of Delaware, Newark, Delaware

A quantitative study has been made of the effect of the major bubble plate design and operating variables upon gas film and liquid film plate efficiency. Since these variables affected the gas and liquid mass-transfer rates differently, the importance of measuring and correlating gas and liquid efficiencies separately is emphasized. Gas and liquid efficiencies were shown to be related to the foam height and foam density on the operating tray, since each of these factors affects the interfacial area and time of contact for the mass-transfer operations. Correlations for foam height and foam density are presented, permitting prediction of the effect of changes in gas and liquid rate, liquid depth, and length of liquid travel upon plate performance. In most cases such predictions agreed with the experimental gas and liquid efficiency results presented for a large and a small tray. The gas and liquid efficiencies of this study, suitably corrected for changes in physical properties, were combined for prediction of plate efficiencies for the methanol-water system at 1 atm. and at 4 atm. total pressure and over the complete concentration range. These predicted values agreed quite well with new experimental values presented in this paper.

PLATE efficiency studies reported in two earlier papers from these laboratories have shown the usefulness of separately determining the amount of mass transfer taking place in the gas phase and in the liquid phase of a bubble tray (*4, 8*). Gas film efficiencies reported in these studies were found by measuring the approach to complete humidification attained by air bubbling through water at the wet bulb temperature; liquid film efficiencies were obtained by determining the degree of desorption of oxygen by air from oxygen-rich water. It was shown that plate efficiencies for C_4 hydrocarbon-furfural-water systems, determined in a companion study, could be predicted satisfactorily from the gas and liquid film efficiency data, by correcting each for the differences in physical properties between the two systems and then combining the two resistances.

Both the single film and the extractive distillation studies were carried out on small trays of identical design, however, and additional studies seemed desirable to determine the effect of important plate design variables upon plate performance. Use of the single film methods seemed especially desirable in studying these variables since a more fundamental analysis of the results could be obtained. For example, plate width is shown to be of great importance in governing liquid efficiency, since the longer the path over which the liquid is contacted by gas bubbles, the greater will be the amount of mass transfer in the liquid phase. On the other hand, in the absence of appreciable hydraulic gradient, plate width does not affect gas efficiency. In this case the gas, in bubbling up through the liquid at a given velocity, contacts the same depth of liquid regardless of plate width. These same concepts are valuable in explaining the effects of liquid depth, skirt clearance, and cap spacing upon plate performance.

An even more fundamental approach to the amount of mass transfer obtainable on a bubble plate, through use of aeration or density factors as a correlating variable, is suggested in this paper. Since these factors are related to the amount of gas entrained in the liquid and, therefore, to the interfacial area, changes in the aeration factor caused by changes in gas rate, cap spacing, and cap size may be directly related to the plate performance.

Since the validity of the single film methods for prediction of overall plate efficiencies for any system has been demonstrated within only one system, a further verification of these methods has been made in this paper by comparison of predicted efficiencies with new, precise plate efficiency data for rectification of methanol-water at 1 atm. and at 4 atm. total pressure.

Additional Gas Film Results

In a previous paper from these laboratories (*4*) gas film efficiency data were presented for humidification of air with water on a 13-in. diameter tray containing 13 caps of 1½-in. diameter spaced in three rows. These runs were made over a range of gas rates up to more than 3 ft./sec. based on the bubbling area of the tray, at liquid rates of 18 and 60 gal./min., at a skirt clearance of ¾ in., and with an outlet weir height of 4 in. In the present study, humidification performance has been measured, for this plate, to include cases of zero skirt clearance with outlet weir heights of 4, 3¼, and 2¼ in. In each of these three cases, efficiencies were determined at liquid rates varying from ½ to 50 gal./min. and superficial gas rates varying up to 2.5 ft./sec. A typical plot of the results is shown in Figure 1, where the number of gas film transfer units, N_G, is shown plotted as a function of gas and liquid rate for the plate design of 3¼-in. outlet weir and zero skirt clearance. It may be noted from Figure 1 that experimental points do scatter somewhat, especially at the high liquid rates; but lines were placed through the

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‡ Present address: The Du Pont Co., Martinsville, Va.

Pt. II of this paper will appear in an early issue.

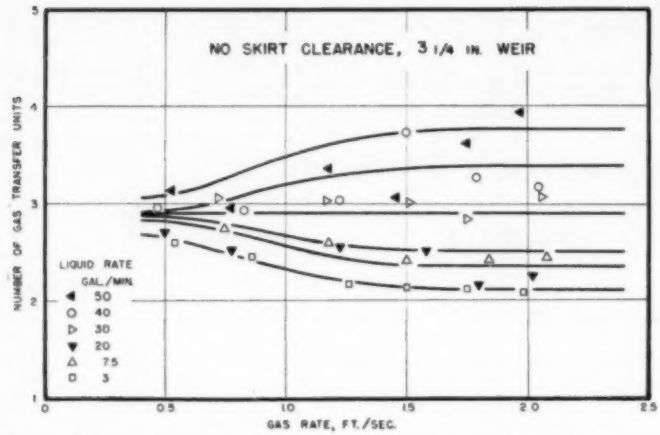

Fig. 1. Typical effect of gas and liquid rate on gas film performance for 13-in. diam. column with 3¼-in. outlet weir and caps flush on the plate.

data points with the help of results obtained with outlet weir heights of 2¼ and 4 in. as well as with the 3¼-in. outlet weir. Value of N_G were computed from E_G, the fractional gas film efficiency, by the relationship (*4*):

$$-2.3 \log (1 - E_G) = N_G \quad (1)$$

The fractional gas film efficiency was calculated to be the ratio of actual change in moisture content of the air stream as it passed through the water on the plate to the maximum change in moisture that would occur if the departing air were fully saturated with moisture. In all cases the water flowed to and from the plate at the wet bulb temperature.

Effect of Liquid Rate upon N_G. Figure 1 shows that as the liquid rate is increased, the gas film efficiency or the number of gas film transfer units on the plate (N_G) increases. This may be explained by the increase in actual foam or froth height on the operating tray which accompanies an increase in liquid rate. Such an increase permits more time for contact of the gas with the liquid, resulting in more mass transfer and greater plate efficiency. If it is assumed that the gas bubbles on the operating tray rise through the liquid at a uniform velocity, the time of contact of the gas with the liquid should be directly proportional to the foam height. Marshall and Pigford (*13*) and Higbie (*10*) have shown that when a liquid and vapor phase are brought into contact, the rate at which material is transferred by unsteady-state diffusion across the interface is proportional to the time of contact raised to the 0.5 power. Thus, N_G values should increase with the 0.5 power of the foam height if these principles apply.

Figure 2 is a log-log plot of N_G as a function of Z_f, the foam height above the tray floor, for two typical superficial gas rates, 0.8 and 1.5 ft./sec. N_G values for liquid rates from 3 to 50 gal./min. and outlet weir heights of 2¼, 3¼, and 4 in. are included in this plot. Although the points scatter somewhat, the data for each gas rate are well correlated by a straight line of slope 0.5.

Consideration was given to the fact that at higher liquid rates the gas bubbles might be carried along horizontally, so that the effective time of contact of the gas with the liquid—and the resulting plate efficiency—would be greater. However, the linear gas rate through the liquid was so much greater than the linear liquid rate that this effect appeared negligible.

Characteristics of Liquid Froth on a Bubble Tray. Since froth or foam height is an important variable in correlating plate efficiencies, methods for predicting foam heights in operating columns seem desirable. For the small tray described above, where the outlet weir was quite close to the last row of caps on the tray, the liquid flowed as froth over the outlet weir. Characteristics of the froth itself were therefore required, and measurements of froth density were carried out with the air-water system at 1 atm. for the small tray. Froth density was computed as the height of equivalent clear liquid on the operating tray, as measured by a static water manometer attached to the tray floor, divided by the total froth height obtained by visual measurement. A slight correction to these results was made in assuming the liquid below the slot opening to be unaerated. Scheibel (*17*) suggests that a more rigorous definition of froth density, ϕ, would be $(\rho_f - \rho_G)/(\rho_L - \rho_G)$ where ρ_f, ρ_G, and ρ_L are densities of the foam, gas, and liquid respectively.

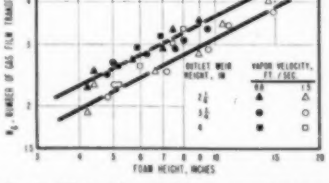

Fig. 2. Effect of foam height on operating tray upon gas film performance. Parameter is gas rate based on superficial bubbling area. Data for outlet weir heights of 2¼, 3¼, and 4 in. are plotted.

Experimental values of froth density, ϕ, are shown plotted in Figure 3 as a function of gas velocity. The values given are the average of many tests made. Some variation of ϕ with liquid rate was found at extremely low and extremely high liquid rates but, in general, ϕ was found to be affected mainly by gas rate. Plotted on the same curve are values of ϕ determined for a large tray section, which also had the last row of caps close to the outlet weir. This tray contained 5 rows of 6¾-in. diam. caps on 9½-in. centers. Efficiency characteristics of this large tray were also determined and are given in detail later. Plotted also on Figure 3 are froth density data by Bagnoli (*1*) for a 13-in. diam. sieve tray containing ⅛-in. diam. holes, spaced either on 1½- or ¾-in. centers. Figure 3 shows a marked difference in froth density between the two bubble trays containing large and small diameter caps. As would be expected, the smaller caps gave a more highly aerated foam, which approached that for the sieve tray at high gas velocities. A fundamental study of factors affecting foam density should prove interesting and worth while. Preliminary studies in our laboratories have shown that foam density for viscous oils is slightly higher than corresponding values for water.

For the tray designs studied, where the liquid flowed as froth over the outlet weir, it was found that the froth flow rate was proportional to the height of froth above the weir raised to the 1.5 power; this is the same relationship valid for unaerated liquids. In Figure 4 experimental values of cubic feet of froth per second per foot of weir length are plotted on log-log paper as a function of the height of froth above the top of the weir. Results are shown for the complete range of gas and liquid rates, and all plate designs employed with the 13-in. diam. column. Data for the large plate section of this study, which has different froth

density characteristics, are also shown on Figure 4. Figure 4 shows that when the froth height is greater than 5 in. above the top of the weir, the froth "collapses" and the 1.5 power relationship no longer holds.

Use of the data in Figures 3 and 4 to predict the effect of changes in gas and liquid rate upon froth height, which in turn affects plate performance—as shown above and later—may be summarized as follows: From the known gas rate, the froth density may be read from Figure 3. The froth rate may be computed by dividing the known clear liquid rate by the froth density. From Figure 4, at the known froth rate, the froth height above the weir may be obtained.

Effect of Gas Rate upon N_G. Figure 1 shows that gas rate has but little effect upon N_G; but this apparently is due to counteracting effects of the many variables involved. One of the effects of an increase in the gas rate is to lower the time of contact of the gas with the liquid, which contributes to a lowering of the amount of mass transfer. To a small degree, an increased gas rate raises the foam height and increases the time of contact. As a further effect, an increase in gas rate increases the degree of aeration of the liquid, which increases the interfacial area between phases and thus increases the amount of mass transfer.

Quantitative treatment of these suggested mechanisms may be shown in the following manner. The relative amount of interfacial area present in a unit depth of foam may be computed by noting that the volume of gas entrained per unit volume of foam is $(1-\phi)$, which is proportional to the volume of gas per unit depth of foam for a given tray. If this gas volume is assumed to be in the form of n spheres it is proportional to nr^3, where r is the radius of each sphere; or, $(1-\phi) = cnr^3$, where c is a constant. The interfacial area is proportional to nr^2, which from the above relationship is proportional to $n^{1/3}(1-\phi)^{2/3}$. Studies on gas bubble formation by van Krevelen and Hoftijzer (*18*) have shown that the number of bubbles formed from a single orifice increases inversely as the volumetric gas rate is raised to the 0.4 power. These results would indicate that $n^{1/3}$ might be a variable which changes but little with gas rate and may be neglected. If this is true, the interfacial area would then be proportional to $(1-\phi)^{2/3}$.

The effect of time of contact of the gas with the liquid may be estimated from the gas rates by considering that the cross-sectional area for gas flow within the foam for each case is different. Assuming, for a given plate, that the area for gas flow within the foam is proportional to $(1-\phi)$, the linear gas rate within the foam is proportional to $u/(1-\phi)$ where u is the linear gas rate based on the open column cross-sectional area available for vapor flow. The time of contact of the gas with the foam is then proportional to $Z_f(1-\phi)/u$.

The amount of mass transferred within the gas bubbles is taken to be directly proportional to the interfacial area and to the time of contact raised to the 0.5 power. This would predict that the effect of gas velocity on N_G could be computed according to its effect on the function $(Z_f/u)^{1/2}(1-\phi)^{7/6}$.

As an example, consider from Figure 1 the effect upon N_G of increasing the gas rate from 1 to 2 ft./sec. when the liquid rate is 7.5 gal./min. Values of froth density, ϕ, for the plate design represented on Figure 1 may be obtained from Figure 3 as 0.44 g./cc. at the lower gas rate and 0.31 g./cc. at the higher gas rate; corresponding values for total froth height, Z_f, from Figure 4 are 4.3 and 4.6 in. The value of $(Z_f/u)^{1/2}(1-\phi)^{7/6}$ at the lower gas rate is 1.06, while at the higher gas rate it is 0.98. Thus the performance at the higher gas rate should be 0.98/1.06 or 0.93 times as great as at the low gas rate; the experimental ratio of N_G values from Figure 1 is 2.3/2.5 or 0.92. A second, similar calculation at a liquid rate of 30 gal./min. predicts a performance ratio of 0.97, compared with an experimental value of 1.0.

Effect of Skirt Clearance and Outlet Weir Height. Values of N_G were experimentally determined for cases of 2¼-, 3¼-, and 4-in. outlet weir heights with zero skirt clearance, and are correlated with previous results for the case of 4-in. outlet weir height and ¾-in. skirt clearance (*4*). Results for any one plate design were similar to Figure 1 and, since gas rate did not affect performance to any extent, values of N_G are plotted in Figure 5 as a function of liquid rate for gas rates of 0.8 and

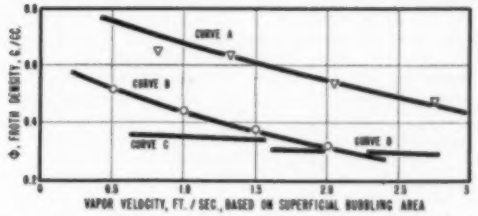

Fig. 3. Froth density on operating bubble tray as a function of vapor velocity. All data for air and water at 25° C.

Curve A: Large tray section with 5 rows of 6⅛-in. diam. caps on 9½-in. centers.
Curve B: 13-in. diam. tray with 3 rows of 1⅝-in. diam. caps on 2½-in. centers.
Curve C: Data of Bagnoli (1) for 13-in. diam. sieve tray containing ⅛-in. holes on 1⅛-in. centers.
Curve D: Data of Bagnoli (1) for 13-in. diam. sieve tray containing ⅛-in. holes on ¾-in. centers.

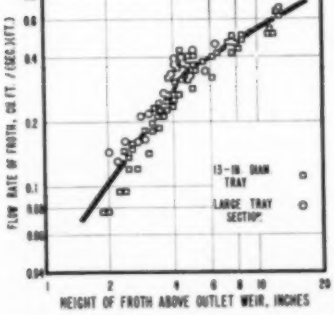

Fig. 4. Correlation of froth height data. Froth rate expressed as (cu. ft. of froth)/(sec.) (ft. of weir length) is plotted as a function of froth height. Data from the 13-in.-diam. tray include results for cases of 3¼- and 4-in. outlet weir heights with no skirt clearance, and 4-in. outlet weir height with ⅜ in. skirt clearance. Data from the large tray section are for the case of 2½-in. outlet weir and ¾-in. skirt clearance.

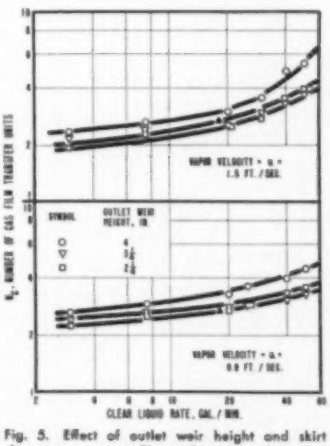

Fig. 5. Effect of outlet weir height and skirt clearance upon gas film performance. Two plots are shown each at a different vapor velocity. Parameter is height of outlet weir in inches. All points are for caps flush on the plate except for points shown as solid triangles, which represent the case of ¾-in. skirt clearance and 4-in. outlet weir height.

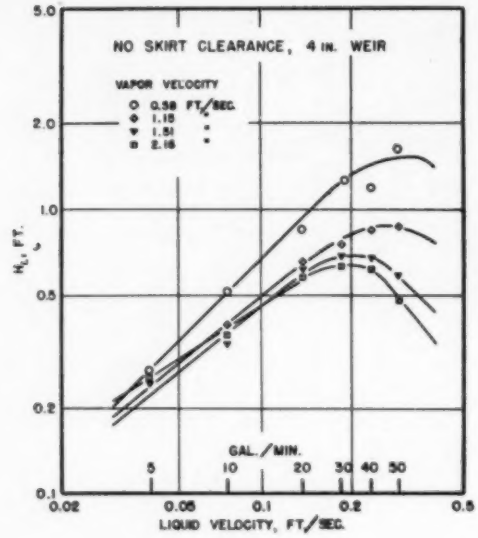

Fig. 6. Typical effect of gas and liquid rate on liquid film performance for 13-in. diam. column with 4-in. outlet weir and caps flush on the plate.

1.5 ft./sec. based on the bubbling area. An increase in outlet weir height is seen in Figure 5 to cause a small but definite increase in performance. This latter result is caused only by the increase in foam height which accompanies the increase in weir height, as shown in Figure 2 where N_G values were affected by foam height raised to the 0.5 power. This may be illustrated by considering the case of increase of outlet weir height from 2¼ to 4 inches at a gas rate of 0.8 ft./sec. and a liquid rate of 20 gal./min. From Figures 3 and 4 the total foam height changes from 4.15 in. to 5.9 in. The corresponding performance at the high weir height should be $(5.9/4.15)^{0.5}$ or 1.19 times as great as at the low weir height. This compares favorably with the ratio of N_G values at these two conditions taken from Figure 5 which is 3.2/2.7 or 1.2. Experimental values at other conditions of gas and liquid rate check equally well.

The effect of change in skirt clearance is also shown in Figure 5 where it may be seen that the performance with a plate design of 3¼-in. outlet weir height and zero skirt clearance is the same as that obtained with a design of 4-in. outlet weir height and ¾-in. skirt clearance. This apparently means that where the caps are raised ¾ in. above the plate, the liquid between the tray floor and the bottom of the cap is not contacted by gas bubbles; or in other words, the effective vertical height of travel of the gas bubbles is the same in both cases. This means that in calculating gas film performance where caps are not flush on the tray, only the active foam height should be considered.

Additional Liquid Film Results

Further liquid film efficiency data in addition to those reported earlier (*4*) have been obtained for the same plate designs used above in the humidification runs. The procedures followed were identical to those used previously (*4*) and consisted of blowing air upwards through the plate across which oxygen-rich water flowed. The degree of stripping of the oxygen from the water compared to the maximum stripping possible was taken as a measure of the liquid efficiency. Because of the extremely low solubility of oxygen in water, effectively all of the resistance to mass transfer was in the liquid phase. The previously published data were obtained with a 4-in. outlet weir and a ¾-in. skirt clearance on the test plate, and the additional data presented below were obtained with zero skirt clearance and outlet weir heights of 4, 3¼, and 2¼ in. In all cases, superficial vapor velocities were varied from 0.6 to 2.2 ft./sec. and liquid rates were varied from 3 to 50 gal./min.

Liquid film efficiency results were first converted to N_L, the number of liquid film transfer units, by the relationship of Equation (1) written for liquid efficiency as $-2.3 \log (1 - E_L) = N_L$. As shown later, the amount of mass transfer in the liquid is a direct function of Z_H, the horizontal length of travel over which the liquid is contacted

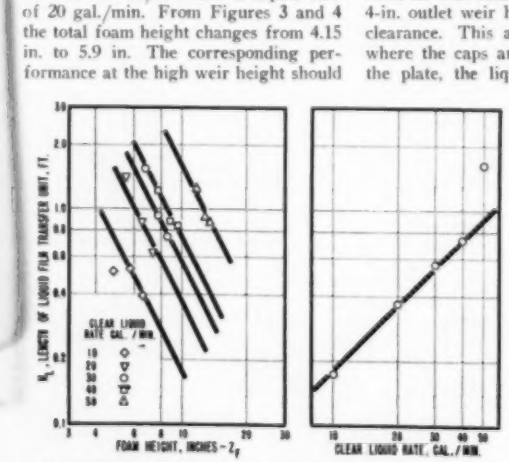

Fig. 7. Effect of change in foam height caused by change in height of outlet weir upon liquid film performance. Curve at left shows values of H_L plotted as function of foam height at constant gas rate of 1.15 ft./sec. (Based on superficial bubbling area) with liquid rate as parameter. The three points defining each liquid-rate curve represent values at outlet weir heights of 2¼, 3¼, and 4 in. Curve at right is cross-plotted from curve at left at constant foam height of 10 in. and shows effect of liquid rate upon H_L in absence of foam height effect.

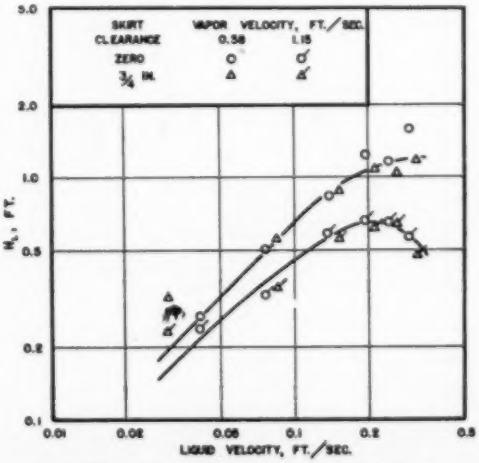

Fig. 8. Effect of skirt clearance on liquid-film performance.

by gas bubbles, so that N_L values were converted to values of H_L, the length of a liquid film transfer unit, by the relationship $H_L = Z_B/N_L$. Z_B for the present case was taken as 8 in.

Typical results are shown plotted in Figure 6 for the case of zero skirt clearance and 4-in. outlet weir. Other plate designs yield results similar in shape to this plot. In this figure performance results expressed as H_L are shown plotted as a function of liquid velocity, with superficial gas rate based on the bubbling area as parameter. Liquid velocity is expressed as maximum linear velocity based on minimum clear liquid flow cross section. As shown later, such a function permits comparison of liquid film data between trays of different designs since it is indicative of the liquid velocity past the slots.

Effect of Gas Rate upon H_L. The effect of a change in gas rate upon liquid phase performance is equally as complicated as its effect upon N_G, but analysis of the problem may be made along similar lines. The relative amount of interfacial area present in a unit thickness of foam moving in a vertical plane normal to the direction of liquid flow is proportional to $(1-\phi)^{2/3}$, as derived earlier for a horizontal plane of unit depth normal to the direction of gas flow; but in this case it is proportional to the foam height, Z_f, as well. Gas rate also affects the time of contact of the liquid with the gas since, by changing the froth density and the foam height, it changes the cross-sectional area available for liquid flow. Quantitatively, for a given plate the true linear velocity of the foam across the plate is proportional to $q/(Z_f\phi)$, where q is the volumetric flow rate of clear liquid in gal./min.; and the time of contact of the liquid with the gas is proportional to the reciprocal of this function. This assumes, of course, that the foam is of uniform density and may be in error for cases of high skirt clearance where the liquid near the tray floor is not aerated. The amount of mass transfer in the liquid phase is shown later by the experimental results to be affected directly by the time of contact of the liquid with the gas rather than according to the time of contact raised to the 0.5 power as predicted by theory (*10, 13*) for the case of unsteady-state diffusion. This relationship is apparently valid for correlating the gas film results since each gas bubble probably remains as an entity during its course of travel permitting a change in concentration within the bubble with time. The liquid on the other hand is well-mixed on the plate and is more or less uniform in concentration with time as it progresses across the plate.

Summarizing, for a given plate operating at a definite liquid rate q, a change in gas rate should affect liquid performance according to its effect upon interfacial area and liquid contact time or according to the change produced in the function $(1-\phi)^{2/3}Z_f(Z_f\phi)$. This simplifies a little to $\phi(1-\phi)^{2/3}Z_f^2$. This relationship may be tested by comparing predicted values with those taken from Figure 6. Consider the effect upon H_L of increasing the gas velocity from 1.15 to 2.16 ft./sec. at a constant clear liquid velocity of 20 gal./min. Values of froth density, ϕ, from Figure 3 are 0.42 g./cc. at the low gas rate and 0.30 g./cc. at the high gas rate; corresponding values for total froth height from Figure 4 are 6.7 and 7.4 in. Performance in terms of H_L at the low gas rate compared to that at the high gas rate may be computed according to the ratio of $\phi(1-\phi)^{2/3}Z_f^2$ values as $(0.30/0.42)(0.70/0.58)^{2/3}(7.4/6.7)^2$ or 0.99; experimental values are in the ratio of 0.64/0.54 or 1.18. A similar calculation at a clear liquid velocity of 40 gal./min. predicts an H_L ratio of 1.52 compared with an experimental ratio of 0.85/0.61 or 1.41; values at 5 gal./min. are 0.81 (predicted) versus 0.92 (experimental). The predicted values deviate around 15% from the experimental values, probably due to the importance of foam height in the calculation which cannot be predicted or even measured too accurately.

Effect of Liquid Rate upon H_L. A change in liquid rate affects plate performance according to its effect upon interfacial area and time of contact of the liquid with the gas bubbles. As shown above, liquid contact time for a given plate is proportional to $(Z_f\phi)/q$; but since ϕ is a function of gas rate alone, a change in liquid rate affects liquid contact time according to the change produced in Z_f/q. Similarly, it was shown that for a given plate the interfacial area was proportional to $(1-\phi)^{2/3}Z_f$, so that a change in liquid rate only affects interfacial area according to the change produced in Z_f. Combining both effects, performance changes caused by a change in liquid rate should be proportional to the change in the function Z_f^2/q.

Known values of H_L, from Figure 6, may be used to test the validity of this proposed relationship for the prediction of the effect of liquid rate. Consider the effect upon performance when the liquid rate is increased from 5 to 40 gal./min. at a constant gas rate of 0.58 ft./sec. The foam heights obtained from Figures 3 and 4 are 4.7 and 6.9 inches at the low and high liquid rates respectively, so that the ratio of the function Z_f^2/q for the two cases is $(4.7/6.9)^2/(5/40)$ or 3.7—indicating that the value of H_L at 40 gal./min. should be 3.7 times greater than the value at 5 gal./min. Actual values of H_L from Figure 6 are 1.17 ft. and 0.27 ft. and their ratio is $1.17/0.27 = 4.3$. Repeating the calculation for the highest gas rate of 2.16 ft./sec., the predicted ratio of H_L values for the same two liquid rates is 2.9, while the experimental H_L values are in the ratio of 2.3.

Effect of Outlet Weir Height on H_L. In the previous section it was shown that interfacial area and liquid contact time on the plate were directly proportional to the function Z_f^2/q, so that for any given liquid rate where q is constant, a change in foam height caused by raising the outlet weir should increase liquid performance according to the change in Z_f^2.

The effect of change in outlet weir height is shown in Figure 7 where experimental values of H_L are plotted for 2¼-, 3¼-, and 4-in. outlet weir heights as a function of the experimental foam heights on the operating tray. The gas rate for all data points was 1.15 ft./sec., and liquid rate variations from 10 to 50 gal./min. are shown as parameters on this plot. It is evident from this figure that the straight lines of slope −2.0 on the log-log plot correlate the data extremely well. Also plotted on Figure 7 are values of H_L taken from these straight lines at a foam height value of 10 in. as a function of q, the liquid rate expressed in gal. of clear liq./min. Except at the highest liquid rate, the performance is seen to vary inversely as q. Thus, at constant gas rate, liquid film performance is shown by Figure 7 to be proportional to Z_f^2/q, as previously derived.

Effect of Skirt Clearance on H_L. Comparison of experimental values of H_L for the case of caps flush with the plate and caps raised ¾ in. above the plate is shown in Figure 8. In both cases the outlet weir height was 4 in. Performance is apparently the same for these two cases over the range of gas and liquid rates investigated. For the case of gas film efficiencies, results obtained with a ¾-in. skirt clearance and a 4-in. outlet weir were more nearly identical with those obtained with zero skirt clearance and 3¼-in. outlet weir. This leads to the conclusion that, for the case of appreciable skirt clearance, the liquid passes freely from the space below the caps to the space above, while the gas seldom penetrates the liquid below the caps. Thus, for the case investigated, raising the caps above the tray floor is detrimental to gas film efficiency, but does not affect liquid film efficiency.

(Acknowledgment, Notation, Literature Cited, and Discussion will run in Part II.)

ABSTRACTS

Abstracts of current articles published in Chemical Engineering Progress Monograph Series, No. 1, Vol. 47, Reaction Kinetics in Chemical Engineering and Chemical Engineering Progress Symposium Series, No. 1, Vol. 47, Ultrasonics—Two Symposia. Volumes are on sale from Chemical Engineering Progress, 120 E. 41st Street, New York 17, N. Y.*

REACTION KINETICS IN CHEMICAL ENGINEERING

O. A. Hougen

University of Wisconsin

A CRITICAL survey of the present status of published literature on reaction rates as applied to process design and an indication of promising fields for imediate development are presented. The seven chapters deal with the scope of applied kinetics, transfer of heat, mass, and momentum; homogeneous reactions, structure and activity of solid catalysts, reactions catalyzed by solid surfaces, uncatalyzed heterogeneous reactions, and empirical methods. Topics of recent engineering development include transient phenomena, fluidized reactions, and temperature gradients.

Chem. Eng. Progress Monograph Series—"Reaction Kinetics," **47** No. 1, 74 pp. (1951).

APPLICATION OF SONIC ENERGY TO COMMERCIAL AEROSOL COLLECTION PROBLEMS

E. P. Neumann and J. L. Norton

Massachusetts Institute of Technology

BASIC factors pertinent to the application of sonic agglomeration on an industrial scale are discussed. Important variables considered are as follows: frequency and intensity of the sound field; time of exposure to the sound; number of particles per unit volume of aerosol; size, shape and other physical properties of particles; degree of agglomeration required. Brief data are given on experimental apparatus, laboratory and pilot plant tests.

Chem. Eng. Progress Symposium Series—"Ultrasonics," **47** No. 1, 4 (1951).

APPLICATIONS OF SONIC ENERGY IN THE PROCESS INDUSTRIES: I. FUNDAMENTALS

C. A. Stokes, Godfrey L. Cabot, Inc.

J. E. Vivian, M.I.T.

BASIC technical facts concerning the applications of sonic energy to the processing of matter to effect physical and/or chemical changes are presented. Fundamental concepts—the nature of sonic energy, its uses and effects—are briefly reviewed. Detailed data are given on: (1) agglomeration of carbon black smoke; specifically, laboratory and pilot plant data to show comparison of sonic with static agglomeration technique; (2) other uses in the gas phase; and (3) liquid phase applications.

Chem. Eng. Progress Symposium Series—"Ultrasonics," **47** No. 1, 11 (1951).

PHYSICO-CHEMICAL EFFECTS OF ULTRASONICS

Alfred Weissler

Naval Research Laboratory

RESULTS are given of investigations of three physico-chemical effects of intense ultrasonic waves in liquids: Depolymerization of polystyrene in toluene and hydroxyethyl cellulose in water, delay in the formation of colloidal sulfur, and liberation of iodine from an aqueous solution of potassium iodide and carbon tetrachloride. Quantitative effects of variations and proportions of reactants are determined. The focussing type of barium titanate transducer is compared with the conventional flat quartz transducer from the standpoint of yield obtained.

Chem. Eng. Progress Symposium Series—"Ultrasonics," **47** No. 1, 23 (1951).

COLLOIDAL EFFECTS OF ULTRASONICS

Karl Sollner

Experimental Biology and Medicine Institute, National Institutes of Health

INDUSTRIAL utilization is considered in the light of phenomena produced: dispersion effects—mutual emulsification of two bulk liquid phases, disintegration of certain solids, peptization, liquefaction of thixotropic gels, depolymerization of high molecular substances, formation of fogs, etc.; and—accumulation, aggregation, coagulation in liquid and gaseous systems, orientation in anisometric particles; crystallization from solution and effect on crystalline structure of metals and alloys.

Chem. Eng. Progress Symposium Series—"Ultrasonics," **47** No. 1, 30 (1951).

ULTRASONICS IN THE STUDY OF MATTER

J. C. Hubbard

Catholic University of America

A GENERAL review of investigations in precision measurements of ultrasonic velocity, the ultrasonic interferometer, and various optical methods such as those based upon the effect of ultrasonic waves as a diffraction medium, Schlieren pictures, and shadowgraphs. Examples for illustration are chosen from velocity measurements in water and mercury, precision work with organic solvents, data on the critical point, and use of pressure and temperature as variables.

Chem. Eng. Progress Symposium Series—"Ultrasonics," **47** No. 1, 41 (1951).

MATERIAL TESTING AND GAGING WITH ULTRASONICS

Benson Carlin

Smith-Meeker Engineering Co.

TECHNIQUES most commonly used are discussed as follows: (1) continuous waves, which are passed through a material; (2) pulse techniques, in which a short bundle of waves or vibrations lasting a few microseconds are transmitted into a material and received after reflection (or transmission), and time of travel measured; (3) resonance effects, in which frequency is varied until resonance in the material occurs. Applications of the first are in specialized cases, of the two latter more widely in metal, aeronautical, and other industries.

Chem. Eng. Progress Symposium Series—"Ultrasonics," **47** No. 1, 46 (1951).

*(See advertisement, page 54, for details.)

CONTINUOUS COUNTERCURRENT ION EXCHANGE

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A continuous ion-exchange system has been constructed in which essentially countercurrent flow was obtained in exhaustion and regeneration columns. In copper-hydrogen exchange, a fluidized exhaustion column may be used because hydraulic classification according to the amount of copper on the resinous exchanger acts to maintain countercurrent action.

Runs were made in which the recovery of copper exceeded 99%. Regeneration with sulfuric acid produced solutions containing up to 6.8% copper by weight. This was a chemical utilization of 73% of the acid used.

Results were correlated by assuming diffusion in the liquid as rate-controlling. The continuous system was found attractive and feasible on a laboratory scale.

WHEN ion exchange is to be applied to the removal of potentially recoverable metals from a waste stream, several factors are introduced which are not met in those simpler applications where the exchanged ion is discarded with the regenerant, as is the case with ordinary water treatment. These factors are:

1. In order to be suitable for subsequent recovery processes the regenerant should contain a high concentration of the metal ion.
2. The regenerant flow rate and composition should be approximately constant to minimize the need for equalizing tanks.
3. Utilization of the regenerant should be nearly complete so that there is little of the unreacted reagent, e.g., sulfuric acid, in the concentrated metal ion stream.

These requirements are difficult to meet in the ordinary cyclical fixed-bed manner of operation, but they should be attainable with continuous countercurrent operation. This method of operation should result also in a smaller amount of resinous exchanger being required for a specific application, with consequent reduction in the necessary investment.

Continuous ion exchange is not a new idea. As early as 1921 Nordell (*8*) applied for patents for continuous water-softening apparatuses using natural zeolites, although these systems apparently did not come into general use. With the advent of the high capacity bead-form resinous exchangers, however, continuous systems became more feasible, because they can be handled by fluidizing in water without great attrition loss. Even so, when the exchanged ion is to be discharged only as waste, the full benefits of continuous exchange probably cannot be realized, although economies in regenerant usage and reduction of investment in equipment would be of interest.

Probably the industrial process most closely analogous to continuous ion exchange is hypersorption (*3*). In this process, contact between fluid and solid is effected in approximately the same manner as required in a continuous ion-exchange system. The fluidized catalytic-cracking process also involves fluid and solid contact and transport, and warrants study in the light of this problem, although it is not ordinarily counterflow. The solvent extraction of oilseed (*4*) also has points of similarity, although cumbersome methods are necessitated by the nature of the seeds.

Theory. Using the treatment of ion-exchange equilibria given by Bauman and Eichorn (*1*), and that of ion-exchange kinetics presented by the authors (*9*), one can develop certain limiting theoretical relationships for continuous ion-exchange systems. In regard to equilibrium, two cases will be considered. First, is that in which the ion exchanged from the solution is held irreversibly by the exchanger. The equilibrium for this case is

$$C^* = 0 \quad \text{for } q < a \tag{1a}$$

$$C^* = C_o \quad \text{for } q = a \tag{1b}$$

where

q = concentration of exchanging ion in the exchanger—millieq./g.
a = total capacity of exchanger—millieq./g.
C = concentration of exchanging ion—millieq./l.
C^* = "back pressure" of exchanging ion—millieq./l.
C_o = total cation concentration—millieq./l.—equal to initial concentration of exchanging ion if solution originally contained no other cation

The authors have shown that this case is approximated in the recovery of copper from dilute solution by a nuclear sulfonic exchanger in the hydrogen or sodium form in fixed beds. In the second case, linear equilibrium prevails.

$$C^* = C_o q/a \tag{2}$$

Equation (*2*) has been shown to approximate the data for the exchange of copper for magnesium as well as the data for the regeneration of copper with sulfuric acid and sodium sulfate in solutions of 10 wt. % and greater concentration. The fact that the equilibrium in systems in which a divalent and a monovalent ion exchange becomes more linear with increasing concentration has been shown to be of considerable importance, as it facilitates the regeneration of beds saturated by a nearly irreversible equilibrium (*2, 9*).

For the first, nonlinear, equilibrium case, the assumption of a liquid diffusion rate mechanism, confirmed by the authors for fixed beds, gives the instantaneous exchange rate, on the basis of a unit weight of exchanger as

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$$dq/d\theta = k_D SC \qquad q < a \qquad (3)$$

$$dq/d\theta = 0 \qquad q = a \qquad (3a)$$

For the linear equilibrium case

$$dq/d\theta = k_D S(C - C_o q/a) \qquad (4)$$

Without specifying the exact methods of mechanical operation one can recognize two types of contact possibly afforded in continuous systems. These are: (a) countercurrent flow, and (b) flow of liquid through a completely mixed bed of solid. Either would be only approximated and not absolutely realized in a practical system. The latter type will be considered first.

Irreversible Exchange Equilibrium. With ionic systems for which the equilibrium approaches the case represented by Equations (1*a* and 1*b*), irreversible equilibrium, the rate of exchange in a continuous mixed reactor depends solely on the amount of exchanger present not yet completely saturated. The distribution of the particles with varying degrees of saturation short of $q = a$ in regard to sequence of contact with the fluid stream does not influence the rate of exchange. Indeed, a reactor in which the particles of exchanger are completely mixed at random would be nearly as effective as a countercurrent system.

Let E = fraction of exchanger present which is not saturated. MacMullin and Weber (7) have shown that this fraction, E, can be expressed as:

$E = 1 - e^{-t R/x}$ E has a value between 0.63 and 1.0 in practical cases

where

t = time for exchanger to saturate

$$= \int_o^a \frac{dq}{k_D SC} \cong \frac{a}{k_D S C_{lm}}$$

in which

$$C_{lm} = \frac{C_o - C}{ln \dfrac{C_o}{C}}$$

x = bed size, g.

R = exchanger feed rate, g./min.

$\dfrac{x}{R}$ = nominal holding time for exchanger

Thus for this reactor with exchanger completely mixed, the exit concentration of the ion which is exchanged irreversibly can be obtained by integration of equations, (3), and the material balance.

$$Rdq = -VdC \qquad (5)$$

$$\frac{C}{C_o} = -e^{\frac{-EkDSx}{V}} \qquad (6)$$

where

V = liquid rate—l./hr.

The case of countercurrent flow may have simpler treatment, and it results in an equation the same as (6) but with E equal to unity. Equation (6) shows that supplying an excess of resin and making the column long enough can give low values of C/C_o with either a countercurrent or a stirred contactor.

Linear Exchange Equilibrium. With linear equilibrium, the mixed contactor is greatly limited in the completeness of removal which it can effect. If, for example, stoichiometric amounts of completely regenerated resin and feed liquid were used, the minimum C/C_o to be obtained from a reactor of infinite volume would be 0.5. Thus for the ionic system with linear equilibrium, or in general any other than completely irreversible equilibrium, countercurrent flow of liquid and resinous exchanger is highly desirable.

With countercurrent systems, for equilibrium cases for which there is a "back pressure" of the exchanging ion, the effluent concentration can be calculated by integrating the suitable rate and material balance equations. Generally, a graphical construction similar to that for a countercurrent gas-absorption tower can be used, but for the special case of linear equilibria, the simple expression

$$V(C_o - C) = k_D S x(\Delta C)_{lm} \qquad (7)$$

in which

$$\Delta C = (C - C^*)$$

and

$$(\Delta C)_{lm} = \frac{\Delta C_{top} - \Delta C_{bottom}}{ln \dfrac{\Delta C_{top}}{\Delta C_{bottom}}}$$

results from formal integration. Equation (6) with $E = 1$ may be converted to this form and thus Equation (7) applies to both linear and irreversible equilibria if operation is counterflow.

Characteristics of Fluidized Beds

Hydraulic Classification. A fluidized ion-exchange bed is one with liquid upflow at a velocity sufficient enough to eliminate the frictional forces between the individual particles. In such a bed, the degree of longitudinal mixing is influenced by differences in diameter or density in the particles in the bed. It would be expected from the principles of elutriation that with particles similar in all properties but diameter, the bed would be classified with the largest at the bottom. This effect has been noticed and exploited in upflow ion-exchange beds (9) of the usual noncontinuous sort. Such a classification action, it will be shown, is undesirable in the case of a continuous system, and hence the use of exchanger particles of nearly uniform diameter is advantageous. With such uniform particles it has been found that the exchanger bed will become classified according to density, that density being a function of the chemical form of the exchanger.

A vertical glass column of 1.1 cm. I.D. was used to study classification. A batch of 24 g. of a 28-35 mesh of Amberlite IR-120 was taken. One half of this quantity had been left in the hydrogen form, the other half had been partially saturated with copper to the extent of 3.1 millieq. Cu/g. and then rinsed with water. The two portions were mixed together and then placed in the column. Distilled water was run into the column from the bottom at a rate to provide 60% expansion for 10 min. The bed was then allowed to settle and the resin was removed as a piston. Samples were taken for analysis. The results are given in Table 1.

TABLE 1.

Bed Height	Depth of Sample Pt. (cm.)	q (millieq. Cu/g.)
36 cm.	0-2	0.26
	11-13	0.27
	20-22	2.47
	31-33	2.77

These results show clearly that with the conditions used, the exchanger particles in a fluidized bed used in copper-hydrogen exchange will be classified with saturated particles at the bottom.

The relative free-settling velocities of exchanger particles in different chemical forms were assumed to be indicative of the direction of hydraulic classification which would take place in different ionic systems.

Determinations of the free-settling velocities of particles of Amberlite *IR*-120 in several chemical forms in water gave results as shown in Table 2.

TABLE 2.

Chemical Form of Exchanger	Equiv. wt. of Ions	Free Settling Velocity, Relative to Hydrogen Form
H	1	1.00
Mg	12	1.07
Na	23	1.17
Cu	31.2	1.28

Thus it can be anticipated that the classification by saturation in a binary system would be such that the particles predominantly in the form with higher equivalent weight are at the bottom. The sharpness of the classification would probably be a function of the magnitude of difference. At best, a fluidized reactor, with solid exchanger entering at the top and leaving at the bottom and liquid flowing upward can be nearly countercurrent in action.

Fluidized and Moving Beds. The ionic systems to be used were copper-hydrogen, copper-sodium, and copper-magnesium, the aim being to develop the operation for the concentrating of dilute copper sulfate waste solutions with most effective use of regenerant.

The selective affinity of the exchanger for copper over hydrogen in dilute solution would permit the use of a mixed contactor with fair degrees of cleanup. However, the hydraulic classification

characteristics of the exchanger would provide roughly countercurrent action in a fluidized bed, if the exchanger went from top to bottom and the liquid from bottom to top. Subsequent work with screen-plates dividing the fluidized bed into a series of countercurrent stages has shown little advantage for copper-hydrogen exchange, although the more positive countercurrent action so provided might be necessary for other ionic systems (*5*).

For the regeneration, the equilibrium is nearly linear (*see* Fig. 1). This makes countercurrent flow mandatory. The hydraulic classification, with copper-saturated particles going to the bottom, would be deleterious. This called for a "moving bed," i.e., one which traveled in piston-like flow with no movement of particles relative to one another. The high concentration of the regenerant solution led, through stoichiometric limitations, to the fact that the regenerant flow rate was low and so would not cause fluidization of the regeneration bed.

If a weak acid resinous exchanger, such as a carboxylic, were employed instead of a nuclear-sulfonic type, the equilibria for the exhaustion and regeneration steps would dictate different requirements for the flows in each leg.

Apparatus and Procedure. The design of the apparatus used follows directly from the theoretical aspects just described above. The exhaustion column was of 13.2 mm. I.D. glass tubing and the regeneration of 8 mm. I.D. tubing. The resinous exchanger was fed to the exhaustion and regeneration columns by a rotating disk feeder. This is shown in Figure 2. Resin in regenerated and saturated forms was held submerged in water in hoppers above the feeder. The regenerated resin entered the left-hand side of the upper plate, fell by gravity into the holes in the outer ring of the rotor, and was discharged to the exhaustion column through the hole in the lower plate. The exhausted resin followed a similar path through the right side and the inner holes. The plate rotation rate range was from 1 to 10 rev./min. By placing rubber plugs in one or two of the holes in each ring of the rotor, the range of feed rates could be altered.

The flow of the solid exchanger from the bottom of each of the columns was controlled by apportioning the liquid feed between the upper and lower feed points. (*See* Fig. 3.) The lifting of the exchanger from the collecting tubes at the bottom to the hoppers above the feeder was not performed continuously in the experimental apparatus, although it can be assumed that this shortcoming could be eliminated in a full-scale design. For example, Crits (*5*) has done this with a continuous air lift. By clamping off the rubber tubing at points A and A' and then opening clamps B, B' and C, C' distilled water was used to force the exchanger through the rubber tubing to the proper hopper. In the 20 or fewer seconds required for this lifting operation, the outgoing exchanger did not back up into the columns. It is felt that column operation was, therefore, essentially continuous.

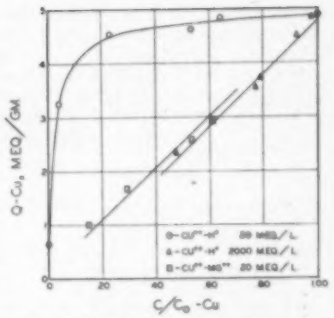

Fig. 1. Equilibrium for copper exchange. Amberlite IR-120.

Making a run consisted of operating the system under constant inlet conditions until steady state was approached. Samples of effluent and regenerant were collected and analyzed for copper colorimetrically (*6*). The rate of collection of the samples gave the liquid effluent rates: the inlet rates

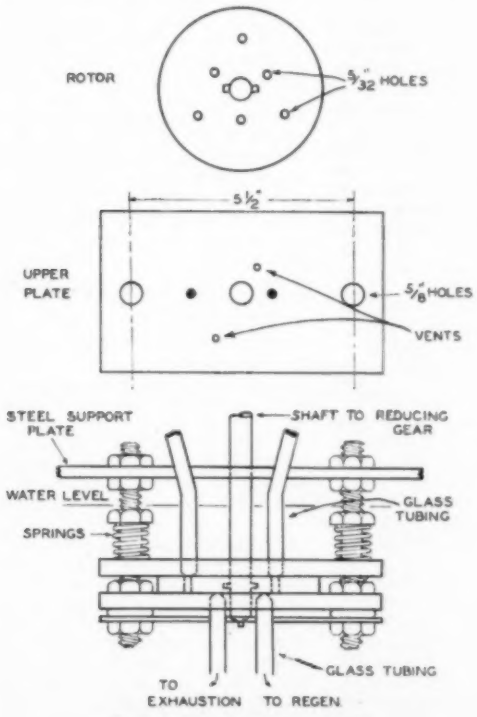

Fig. 2. Submerged rotating resin feeder.

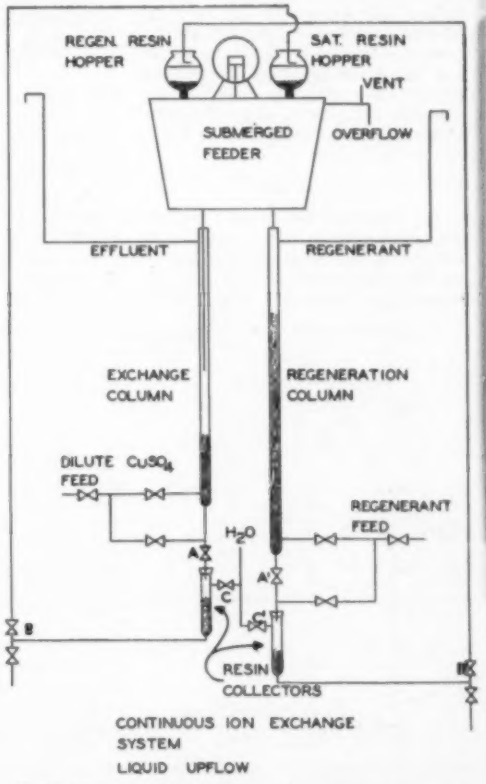

Fig. 3. Schematic view of continuous cation concentrating system.

TABLE 3 a.- Exhaustion leg results.

Run No.	25	26	30	31	35	40	41	42	50	51	54	300	301
Bed Ht.-Cm	19.4	24	20	20	18	10	10	20	24	24	24	20	20
-G.	5.8	6.9	5.5	5.5	5.4	3.4	3.1	6.4	6.4	6.9	6.9	5.8	6.7
C_o-m.eq/l.	20	20	20	20	20	20	20	20	5.2	5.2	21	20	20
Ionic system with Cu	H	H	Mg	Mg	Na	H	H	H	H	H	H	Na	Na
V - L/hr.	2.3	2.7	2.7	2.7	2.4	1.5	2.3	2.3	3.0	3.0	1.5	2.5	1.6
R g./hr.	24	22	19.4	28	38	9	1.5	15.6	8.4	3.7	9	12	12
Q-entering, m.eq/g.	0.02	0.02	0.02	0.02	0.02	1.10	1.2	0.03	0.32	0.32	0.07	0.80	0.80
C/C_o	0.01	0.02	0.045	0.042	0.042	0.016	0.14	0.030	0.009	0.01	0.007	0.04	0.007

TABLE 3 b.- Regeneration leg results.

Run No.	101	103	104	91	92	93	94	95	96	98
Bed Ht.-Cm	25	50	50	50	50	50	50	67	72	70
Wt.-G.	4.5	8.2	9.3	9.2	10.2	9.5	10.4	13.5	13.5	13.4
Regenerant ion	Na	Na	Na	H	H	H	H	H	H	H
Normality of regenerant	2.2	2.2	2.2	3.0	3.0	3.0	3.0	3.0	3.0	3.0
V - L/hr.	0.14	0.18	0.11	0.12	0.054	0.084	0.048	0.054	0.084	0.084
R g./hr.	8.6	12.5	16	33	29	34	18	18	24	3.3
Q-entering, m.eq/g.	4.6	4.7	4.7	5.5	3.6	5.6	3.7	3.7	2.7	4.5
Q-leaving	0.22	0.03	0.06	0.44	0.97	1.06	0.37	0.30	0.68	0.01
Cu in regenerant-wt	1.1	1.4	3.4	3.1	6.8	4.7	4.5	4.4	3.8	1.2
Regenerant utilization	10	20	52	34	73	52	49	48	42	13

were indicated by capillary flow meters. Samples of the resinous exchanger were taken from the circulatory stream and analyzed for copper. The exchanger circulation rate was determined by measuring the time of fall of the exchanger level in the hoppers.

Results

Separate results are listed for the exhaustion and regeneration legs of the system. The operation of one leg is related to the other only through the degree of saturation of the exchanger in and out of the leg. Thus the separate listings allow extension of the results to predict the operation of a wide variety of combinations. For results, see Tables 3*a* and 3*b*.

Comment

The efficacy of the complete continuous systems can be seen by considering runs in the exhaustion and regeneration results which match in respect to exhausted and regenerated resin analyses. Such a combination is given below

Exhaustion run	51
Regeneration run	95
Exhaustion: C_o-millieq./l.	5.2
X-bed height, cm.	24
V-l./hr.	3.0
C/C_o	0.01
Regeneration: Bed height, cm.	67
Concentration of Cu	4.6%
Utilization of regenerant	46%
q-leaving exhaustion	4.3
q-regen. millieq./g.	0.36
Concentration factor $\left(\frac{\text{Cu in regeneration}}{\text{Cu in feed}}\right)$	360

The experimental apparatus employed had several shortcomings which need not appear in a larger-scale adaptation. The rotary feeder employed did not provide a tight liquid seal for the tops of the columns and, to avoid excessive leakage of liquid in or out at these points, required careful balancing of the hydraulic heads involved by adjustment of the height of the column effluent lines.

The rinsing of the regenerant from the exchanger was accomplished in the collectors at the bottom of the column and in the lifting and feeding operations. While this was probably adequate, the rinsing step must be given attention. It has been shown that diffusion in the solid exchanger particles sets a minimum time of about 5 min. for rinsing sulfuric acid from Amberlite IR-120 (*5*). This rinsing could be effected in the top of the exhaustion column, but would result in the loss in effectiveness of that portion of the bed for ion exchange.

The exchanger particles descending into the collectors displaced water up into the column. In the exhaustion column, this represented a negligible amount of water. In the regeneration column, the dilution by the displaced water should be allowed for in the choosing of the regenerant feed concentration.

Channeling was not noticed in the apparatus used, although it might be troublesome in large diameter beds.

Interpretation

Results were interpreted by the calculation of liquid mass-transfer coefficients. For countercurrent flow with any equilibrium, the following can be written

$$\int_{C_o}^{C} \frac{-dC}{(C-C^*)} = \frac{k_D S}{V} \int_{o}^{x} dx \qquad (8)$$

For convenience, numerator and denominator of the right-hand side can be divided by C_o. This is shown for Run 25 on Figures 4 and 5. The (q/a) vs. C/C_o plot of Figure 4 is analogous to the x-y diagram used in distillation and

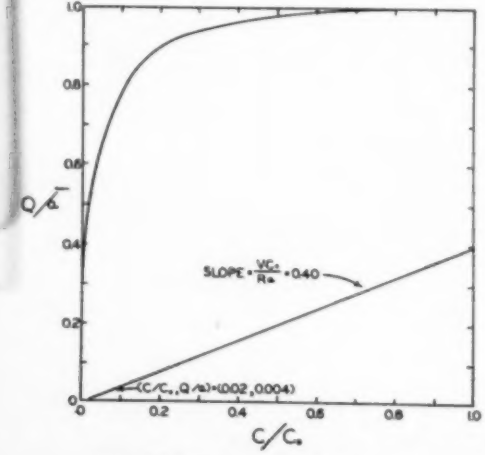

Fig. 4. Operating and equilibrium lines for run 25.

TABLE 4

CALCULATED VALUES OF MASS TRANSFER COEFFICIENT, $k_D S$

Run No.	Superficial liq.vel. L/(hr)(sq.cm.)	Ionic System (if not Cu-H)	$k_D S$ m.eq. / (hr)(g.)(m.eq/L.)
Exhaustion Leg			
25	1.7		1.8
26	2.0		1.5
30	2.0		0.54
31	2.0	Cu-Mg	0.54
35	1.8	Cu-Mg	1.8
40	1.1	Cu-Na	0.9
41	1.7		1.5
42	1.7		1.2
50	2.2		2.2
51	2.2		2.0
54	1.1		0.93
300	1.8	Cu-Na	1.4
301	1.2	Cu-Na	1.2
Regeneration Leg			
91	0.24		0.020
92	0.11		0.018
93	0.17		0.018
94	0.09		0.016
95	0.11		0.012
96	0.17		0.018
98	0.17		0.006
101	0.28	Na-Cu	0.020
103	0.36	Na-Cu	0.027
104	0.22	Na-Cu	0.044

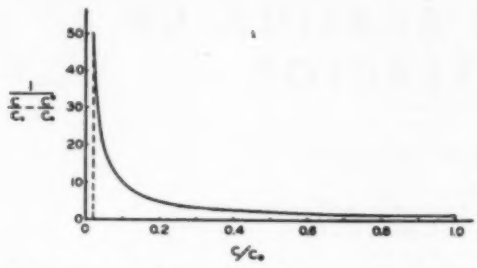

Fig. 5. Graphical integration for determination of rate constant, run 25.

absorption calculations. The operating line represents the material balance Equation (5), and the horizontal distance between the equilibrium and operating lines is proportional to the liquid film driving force for the exchange. In Figure 5 the graphical integration of

$$\int_{C_o}^{C} \frac{d(C/C_o)}{\frac{C}{C_o} - \frac{C^*}{C_o}}$$

is shown. This graphical integration method, or, when applicable, Equation (7), which employs the log mean driving force, gave the values of k_DS. These values appear in Table 4.

The values of k_DS for the exhaustion runs for a given superficial liquid velocity are about one-half those found in fixed beds (9). This is true of the values for each of the three ionic systems studied. The values might be expected to be lower in a fluidized than a fixed bed because the interstitial velocity, which is probably the quantity which actually affects the mass transfer, is greater in the latter case. The dependence of k_DS on velocity cannot be determined with precision from the results presented. In a fixed bed, this has been found to be as the 0.7 power of the liquid velocity. The relationship for a fluidized bed would be expected to be different, because of the increase in fraction of interstitial area with increase in velocity. Over the velocity range in question, for the copper-hydrogen continuous exhaustion runs, the results are in adequate agreement with the simple linear relationship $k_DS = 0.8$ (V/column area). The great difference between exhaustion and regeneration runs probably is the result of the difference in concentration level as much as the greatly different ranges of liquid velocities.

An explanation is possible for the lowest values of k_DS in each set. These correspond to exhaustion runs in which the exchanger left nearly saturated and regeneration runs in which the copper was almost completely removed from the exchanger. When this is the case, the equilibrium imposes severe limitations. In most of the exhaustion operations the equilibrium dictates nearly irreversible removal of copper. At high degrees of saturation, this is no longer true. Even slight longitudinal mixing of the exchanger at the exit end of the bed would have a deleterious effect and would tend to decrease the effective, although not the actual, mass-transfer coefficient. Slight experimental error would lead to considerable deviation in the calculated value of the coefficient.

Comparison with Fixed Beds. A rough comparison can be made between the amount of ion-exchange resin required with fixed bed and with continuous operation. The case to be considered is the concentration of copper from 1,000,000 gal./day of a dilute solution containing 10 millieq./l. of cupric ion as sulfate. The leakage from a fixed bed is to be less than 5%.

Fixed Bed. It has been shown previously that with a *completely regenerated* fixed bed containing 5000 lb. of Amberlite IR-120, the leakage value, C/C_o reaches 0.05 in 5.83 hr. (9). However, in any operation with the usual level of regeneration, a bed containing three times this much resin would be required for this performance. Use of a shorter cycle time would, to a certain extent, allow use of smaller beds, but it is more likely that larger beds would be used with one regeneration per shift. Thus the resin requirement can be taken as two beds each of 5000 lb. or 10,000 lb.

Continuous System.

Exhaustion. Flow = 158,000 l./hr. For flow rate of 2.27 l./(hr.)(sq.cm.) require 70,000 sq.cm. $k_DS = 2.0$ l./(g.)(hr.). For average leakage to be equivalent, C/C_o from continuous system must be less than peak value from fixed bed. Therefore design for $C/C_o = 0.02$. By Equation (8) X = 350 kg. or 770 lb.

Regeneration. Assume Q—entering = 4.20 millieq./g.; Q—leaving = 0.30 millieq./g. For 70% utilization of regenerant—16% H_2SO_4. By Equation (7) with $k_DS = 0.016$; bed = 170 kg. or 375 lb. Allowing for 30% holdup in circulating system—total resin required = 1490 lb.

Some factor of safety should be included in the continuous bed design, but this will be less than the factor of about 3 required on the fixed-bed size necessary for operation with the low regeneration levels necessary for reasonable regenerant efficiency.

Notation

a = ultimate exchange capacity—millieq./g. dry exchanger
C = concentration of exchanging ion in liquid—millieq./l.
C^* = equivalent concentration, "back-pressure" from exchanger—millieq./l.
C_o = initial concentration of exchanging ion, which is equal to total of exchanging and exchanged ion—millieq./l.
E = fraction of number of exchanger particles which are not saturated
k_DS = mass-transfer constant area product, on weight basis millieq./(g.)(hr.)(millieq./l.)
q = concentration of exchanging ion in exchanger millieq./g. dry exchanger
R = resin rate—g. dry exchanger/hr.
t = time for exchanger to saturate, hr.
V = liquid rate—l./hr.
x = bed height—g. of dry exchanger. All exchanger weights based on hydrogen form of resin.
θ = time—hr.

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A PROCEDURE FOR SCALING-UP A CATALYTIC REACTOR

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A procedure is presented for scaling up fixed-bed catalytic reactors. This paper demonstrates the treatment of laboratory data for the preparation of working charts for the designer of the pilot plant and indicates how experience obtained in operating the pilot plant is applied in making the final design. The example used is the production of toluene by the dehydrogenation of methylcyclohexane. A mixture of hydrocarbons—paraffins, methylcyclohexane, and toluene—and recycle hydrogen preheated to 860° F. is passed through tubes containing catalyst pellets. The tubes are heated on the outside by flue gas at 1200° F. Variables considered were: tube size, ratio of hydrogen to hydrocarbons, and toluene in the feed. Final working charts are in terms of volumetric production rate vs. conversion for various tube sizes, recycle ratios and feed compositions, and tube lengths vs. space velocity for various pressure drops.

PROPER pilot plant work stands midway between bench-scale laboratory work and commercial design. It is important that the laboratory experiments supply all pertinent fundamental data for any possible arrangement that the design engineer may suggest. The pilot plant designer should endeavor to anticipate all arrangements of equipment that might arise as a result of the more complete information which will become available as the pilot plant work progresses. In the ideal case pilot plant operation will explore the commercial potentialities of the process and eventually may, depending upon the nature of the reactions, supply operating data on a prototype for the commercial design. In fact, the design of the pilot plant should hinge upon a tentative design for the full-scale plant. With the data obtained from the pilot plant, the engineer who makes the final design of the commercial plant will have, in concise form, information which will permit him to integrate the process in question with other plant processes in the most profitable way. Further development of the process ordinarily will occur even after the commercial plant is put into operation.

Thus an ideal pilot plant starts with the firmly grounded fundamental data from the laboratory and borrows from the body of engineering correlations in order to guess at the probable operation of a large-scale piece of equipment. The work of the pilot plant is, in a sense, to test the application of these correlations. Another duty of the pilot plant is to suggest to the engineer other arrangements which he might not have thought of without this practical application of the laboratory data.

It is apparent that careful planning of the program is essential. Naturally, in many instances the work will not progress chronologically as it was outlined in principle. Press of events may demand concurrent operation of the laboratory and pilot plant work. Pilot plant operation may point out inadequacies in the fundamental data. In many cases the personnel may not be as separate as has been indicated. One man may be the laboratory worker, pilot plant designer and operator, and the designer of the commercial equipment. The pilot plant may even turn out to be the commercial scale plant. Regardless of these inconsistencies, it will be beneficial always to segregate the duties of each phase of development. A common pitfall in the development of a process is to mix these duties. It is usually quite wasteful to determine fundamental information from pilot scale equipment. It is also often the case that inaccurate information will be obtained because of difficulty in interpreting the data. It is just as bad, and for the same reasons, to do pilot plant work on laboratory scale equipment.

This paper illustrates the design of a pilot plant from laboratory data. Because the design of a commercial scale plant is quite similar—since the pilot plant serves only to sharpen one's picture of the engineering correlations—this paper serves also to illustrate the final design.

The problem of designing a commercial reactor on the basis of laboratory measurements has certain features in common with the solution of a complicated differential equation. An ideal tool for the study of a reaction in the laboratory is the differential reactor, which is operated isothermally with a low conversion of all reactants and may be used over a range of all the important variables. A set of measurements made with such a reactor furnishes a tabular or graphical description of the dependence of a reaction rate on the variables concerned, and so, in effect, constitutes a differential equation. The commercial reactor, on the other hand, has large temperature gradients, nearly always is run at a large conversion of the principal reactant, and may have a considerable pressure drop. The problem is to account for the effects of reaction rate, mass transfer, and heat transfer as they vary through the commercial reactor, and to predict the course of the reaction on the basis of the laboratory results. The commercial design may be said to represent an integration of the results of laboratory work with the conditions imposed by engineering considerations.

Several types of data are required before working tables and charts can be prepared. The work in the laboratory should provide data on the rates of all significant reactions as functions of temperature, pressure, concentration of reactants, and the size of the catalyst pellets. The heats of all reactions must be known, and the free energy of all reversible ones. Finally, the effect of the process history—temperature, pressure, and concentration—on the activity and selectivity of the catalyst should be known.

These data must be processed for use by the design engineer. He needs to know how the conversion, production rate, and pressure drop depend on the composition of the feed, the size and shape of the reactor, the pressure, temperature, and flow rate, and the size of the catalyst pellets. In many cases the required quantities need to be calculated for a range of conditions relating to the design of a particular type of reactor; for example, in a fixed-bed tubular reactor the quantity determined is usually the nature of the heat-transfer medium

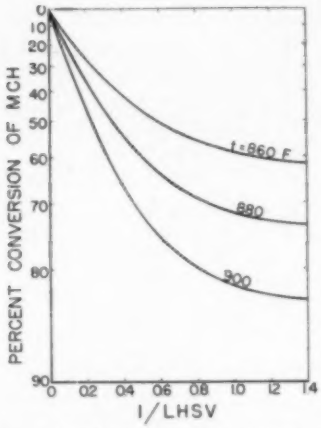

Fig. 1. Dependence of conversion of MCH on flow rate and temperature.

and the temperature at which it enters and leaves the reactor, rather than the temperature of the reactant fluid itself.

The design engineer fixes on the type of reactor to be used—adiabatic, fixed-bed heat-exchanger, fluid-bed, etc. In all cases he must choose the processes to be used for the preparation of the feed and the separation of the products of the reaction. If an adiabatic reactor is to be used, the controlling quantities are the temperature and composition of the feed, the pressure in the reactor, and its volume. On the basis of these quantities, the proper construction of the reactor can be planned. If the reactor is to be of the fixed-bed tubular type, the number, diameter, and length of the tubes must be known, together with the means of transferring heat to or from them and the size of catalyst pellets to be used. If a fluid-bed reactor is to be used, the engineer must decide on the means of heating or cooling it, the size distribution of the catalyst, the flow rate for the reactants, and the size and shape of the reactor. These decisions are based on the way in which the profitability and ease of operation of the process depend on the process variables. This dependence is determined from working tables and graphs derived from data in the literature or obtained from laboratory work. It is only in the ideal case, of course, that the design engineer is furnished all at once with the data he needs to make all these decisions. Usually, he will have to go back to the laboratory worker and to the fundamental chemical engineer for the answers to questions that they had not foreseen.

The procedure that we have outlined will be illustrated in its application to a process for manufacturing toluene from methylcyclohexane (MCH). The dehydrogenation of methylcyclohexane may be carried out catalytically in the presence of other hydrocarbons boiling near MCH. The reaction is reversible in the temperature range that can be used, and at high temperatures there is a significant amount of cracking of MCH and the accompanying paraffins. The laboratory work on the reaction was done with a tubular reactor of moderate size, and since this type of reactor was found to be entirely satisfactory for the process it was adopted for the commercial reactor.

Both the main reaction and the side reactions are endothermic and so favored by high temperature. Desirable conditions for carrying out this reaction would be:

High temperature for high rate and favorable equilibrium
Low pressure for favorable equilibrium
High pressure for fast forward reaction
Small tubes for good heat transfer
Small catalyst pellets for fast reaction
Large catalyst pellets for good heat transfer

Basic data for the main reaction are known, and as no large amount of the side reactions can be tolerated, no new measurements of heat or free energy of reaction were required. The heat of the dehydrogenation of MCH is 94 M B.t.u./lb.-mole (*4*), and the equilibrium constant is about 5.3×10^5 atm.^{3} at 900° F. The effective thermal conductivity and the diffusivity within the reactor were calculated assuming a constant effective Peclét number of 4.4. The over-all heat-transfer coefficient for the transfer between the flue gas and the reactant fluid was taken to be 8 B.t.u./(sq.ft.)(hr.)(° F.).

In the laboratory the effects of space velocity and concentration of toluene in the feed on the conversion of MCH were studied for different temperatures at several pressures. It was found early in the investigation that it was necessary to maintain a high concentration of hydrogen in the reactor in order to avoid severe cracking and rapid loss of the activity of the catalyst. The dependence of the conversion on the concentration of hydrogen was determined. Figure 1 shows the percentage conversion of MCH as it depends on liquid hourly space velocity for three temperatures, all these experiments being made with a molal ratio of hydrogen to hydrocarbon of 7, and at a pressure of 50 atm. These results are adequately described by the rate equation

$$R = k(\mathrm{MCH})\left[1 - \frac{(\mathrm{H_2})^3(\mathrm{tol})}{K(\mathrm{MCH})}\right] \qquad (1)$$

where

R is the rate in lb. mole/(cu.ft.)(hr.)
k is the specific rate constant with dimensions of lb. mole /(cu.ft.)(hr.)(atm.)
K is the equilibrium constant referred to above, and (MCH)(tol) and (H_2) are the partial pressures of MCH, toluene, and hydrogen, respectively.

This means that the forward reaction is first order in MCH while the reverse reaction is fourth order, third in hydrogen, and first in toluene. The apparent activation energy of the forward reaction is 55 M B.t.u./lb.-mole, and the specific rate constant is 10 hr.$^{-1}$ at 880° F.

Other observations made in the course of the laboratory work were that the rate of cracking increases as the temperature is increased and as the partial pressure of hydrogen is decreased, and that a lower partial pressure of hydrogen decreases the effective life of the

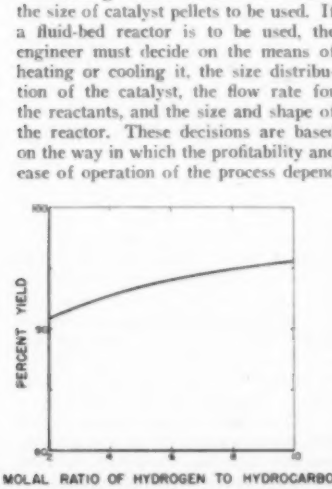

Fig. 3. Effect of partial pressure of hydrogen on liquid yield.

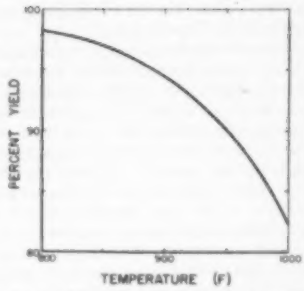

Fig. 2. Effect of temperature on liquid yield.

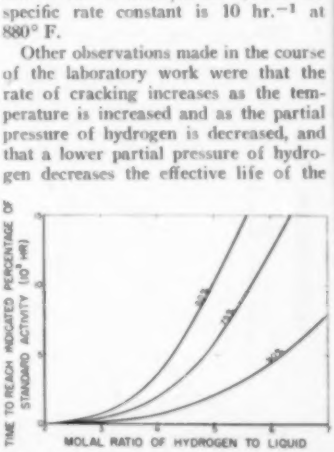

Fig. 4. Effect of partial pressure of hydrogen on catalyst life.

catalyst. These effects are shown quantitatively in Figures 2-4.

Certain features of the commercial reactor were fixed by these last-mentioned effects and by the requirement that the tubes should not rupture at the highest temperature that could conceivably be reached. The requirement of high partial pressure of hydrogen led to a choice of a pressure in the neighborhood of 50 atm. In order to put a firm upper limit on the temperature, no radiant section was included in the reactor, but provision was made to circulate flue gas with an entering temperature of 1200° F. from a separate furnace around the reactor tubes. It was decided to make flow of flue gas and reactants concurrent in order to supply as much heat as possible to the reactants in the region where the reaction would be normally fast. The ratio of flue gas to reactant was chosen so as to give a maximum temperature for the reactants of about 890° F. in order to avoid the cracking associated with excessively high temperatures.

The differential equations governing the relation between conversion, temperature, and reaction rate, as influenced by diffusion and heat conductivity, have been given by several authors (*3*). If simplifications are made, one is left with two second-order partial differential equations, one indicating the conservation of enthalpy and the other the conversion of matter. A good graphical method for integrating these equations has been worked out by Baron (*1*). On starting to use this method, it was found that so large a fraction of the resistance to heat flow was ascribed to the film on the outside of the reactor tubes that the equations could be simplified much further without significant loss of accuracy. If the assumptions are introduced that the concentration and the temperature are quadratic functions of the radial position and that the average rate of reaction is equal to the rate at the average temperature, equations can be put in the form of a first-order ordinary differential equation relating the temperature and the conversion.

An enthalpy balance on a cross section of the bed gives:

$$\frac{\pi D^2}{4} FC_p dt = \frac{\pi D^2}{4} FN_m^\circ(-\Delta H)dx + dq \quad (2)$$

where dq is the heat flow into the bed through the walls. It may be shown that:

$$dq = \frac{U}{1+\frac{DU}{8k_e}}(t_f - t)\pi D dz \quad (3)$$

The factor $(1+DU/8k_e)$ enters the equation to correct the over-all heat-transfer coefficient, U, by taking account of the fact that the difference in temperature between the flue gas and the reactant fluid next to the wall is less than the difference between the temperature of the flue gas and the average temperature of the reactant fluid in a cross section. Insertion of (3) in (2) gives a differential equation with three variables, t, x, z. Since $(t_f - t)$ is a function of x and t, it is best to eliminate the variable, z, by the following relation:

$$aRdz = FN_m^\circ dx \quad (4)$$

Substituting (3) and (4) in (2) we get

$$\frac{dt}{dx} = -\frac{(\Delta H)N_m^\circ}{C_p} + \frac{4UN_m^\circ(t_f - t)}{DC_p aR\left(1+\frac{DU}{8k_e}\right)} \quad (5)$$

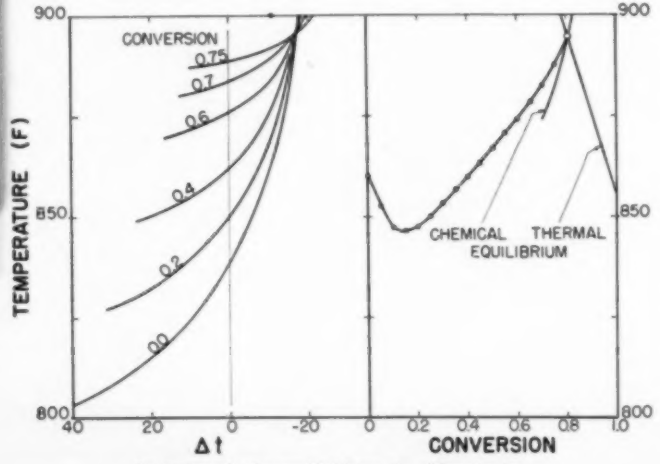

Fig. 5. Example of a graphical integration of Equation (5).

where

F = molal mass velocity of the feed
t = temperature of the reactant fluid
x = fractional conversion of MCH
z = length of the tube
ΔH = heat of the reaction
N_m° = mole fraction of MCH in the feed
C_p = molal heat capacity of the feed
U = over-all heat-transfer coefficient between the flue gas and reactant fluid
t_f = temperature of the flue gas
D = diameter of reactor tube
k_e = effective thermal conductivity in catalyst bed
a = a factor introduced to represent changes in catalyst activity. This factor is taken to be unity for the catalyst used in the experiments shown in Figure 1.

The difference $t_f - t$ can be calculated from a heat balance on the flue gas and the fluid in the reactor, and turns out to be a linear function of the conversion and the temperature.

It is a convenient property of the differential equation relating temperature and conversion that the quantities, D, a, U, and k_e, appear only in the group, $U/Da\ (1+DU/8k_e)$. Thus the relation between temperature and conversion is determined by the value of this group and not by the variables making it up. One may consequently solve the differential equation for a series of values of the group, and then interpolate in this series to get a result corresponding to a specified set of conditions instead of having to solve the equation for a large number of combinations of conditions.

Figure 5 shows the graphical method used for the solution of Equation (5) for the conditions: liquid feed containing 7.5% v toluene; tube diameter, 4.5 in.; mole ratio of hydrogen to liquid, 5. Equation (5) was changed from the differential form to the form:

$$\Delta t = -\frac{(\Delta H)N_m^\circ}{C_p}\Delta x + \frac{4UN_m^\circ(t_f - t)}{DC_p aR\left(1+\frac{DU}{8k_e}\right)}\Delta x \quad (6)$$

The right-hand side of Equation (6) is a function of t, x, and Δx. Δx was chosen as 0.05. The curves on the left in Figure 5 are plots of the right-hand side of Equation (6) as a function of t with conversion as parameter. One uses them by assuming an average temperature for the conversion interval, reading a Δt at the estimated temperature and the average conversion in the interval, and then checking the estimated

average temperature against the observed Δt. This procedure is continued to as close to equilibrium as is desirable. The equilibrium temperature and conversion are determined by simultaneous solution of the chemical and thermal equilibrium equations. Figure 6 shows the approximate relation between temperature and radial position at the point where the average temperature is a minimum in a 6-in. tube, other conditions being like those for Figure 5. The point of minimum temperature was chosen for this illustration because it represents the point at which the difference in temperature between flue gas and the reactant fluid is near a maximum.

The length of reactor required for a given conversion may be found by using Equation (4) in conjunction with the relation between temperature and conversion obtained from Equation (6). If this equation is integrated in the form

$$z = \frac{N_m{}^\circ F}{a} \int_o^x \frac{dx}{R} \qquad (7)$$

and it is noted that the space velocity S is equal to F/z, it may be seen that the space velocity is given by the expression

$$S = \frac{a}{N_m{}^\circ \int_o^x \frac{dx}{R}} \qquad (8)$$

The solution of Equation (6) gives t as a function of x. Since R is known as a function of t and x (See Equation (1) and following), the integral in Equation (8) may be evaluated by a numerical or graphical quadrature.

The relation between temperature and conversion was worked out under several conditions for a feed containing 42% v MCH, 7.5% v toluene, with the remainder C_6 and C_7 paraffins, reacting over a catalyst with standard activity. With a molal ratio of hydrogen to liquid feed equal to 5, calculations were made for tubes, 3, 4.5, and 6 in. in diam.; and with the mole ratio 2, calculations were made for tubes 3 and 4.5 in. in diam. Although for the purpose of illustration the results are shown with all properties except the diameter of the tube kept constant, variations in heat-transfer coefficient, effective thermal conductivity, or activity of catalyst may be represented in terms of equivalent variations in tube diameter. It should be noted that the activity of the catalyst affects the relationship between space velocity and conversion in a fashion different from the other parameters in that it appears both in the group, $U/Da(1 + UD/8k_e)$, and independently in Equation (8). In order to show the effect

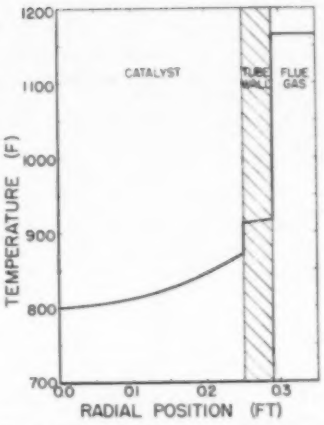

Fig. 6. Calculated radial temperature profile in a 6-in. tube.

of better prefractionation of the feed, the results for a tube 4.5 in. in diam. were calculated for feed like that given above, but with all the toluene removed, using a molal ratio of hydrogen to liquid equal to 5.

Results of the calculations in the accompanying figures show the relation between production rate, that is, average rate of reaction, and the conversion of MCH for the different diameters of tubes, recycle ratios, and feeds. Since the production rate is equal to the product, $SxN_m{}^\circ$, the space velocity, S, can be calculated for a given conversion from the production rate. (See Figures 7 and 8). Figure 9 is an auxiliary drawing showing the relation between space velocity and length of tube for various values of the pressure drop. The length is plotted against the product of the space velocity and the square root of the average molecular weight of the feed including hydrogen for several values of the pressure drop. The factor involving the molecular weight of the feed must be taken into account because the pressure drop depends on the density of the fluid that is flowing. From such drawings the design engineer can find all the relations he needs connecting flow rates, conversions, production rate, and pressure drop.

As an illustration, one can consider a case in which 15 lb.-mole/hr. of toluene are to be produced from a feed which has the following composition:

MCH 42%v
Toluene 7.5%v
Paraffin (mol. wt. = 107) 50.5%v

Other conditions are:

Molal ratio H_2: Liquid = 5
D = 0.375 ft. (4.5 in.)
x = 0.65
Δp = 10 lb./sq.in.

Under these conditions 3.8 moles of toluene are introduced in the feed for each 15 moles produced from MCH. The average molecular weight of the feed stream is 18.5 and the mole fraction MCH is 0.076.

From Figure 7 it is noted that the production rate, $SxN_m{}^\circ$, in the 4.5-in. tube with a conversion of 0.65 is 0.150 lb.-mole/(cu.ft.)(hr.), so that the space velocity, S, is 3.05 lb.-mole/(cu.ft.)(hr.). From Figure 9 one notes that the length of reactor giving a pressure drop of 10 lb./sq.in. is 24.2 ft. Since the total rate of production can be expressed as

$$FxN_m{}^\circ(\pi D^2/4)n = 15$$

where n is the number of tubes, the data needed are now available to determine the number of tubes. It is found

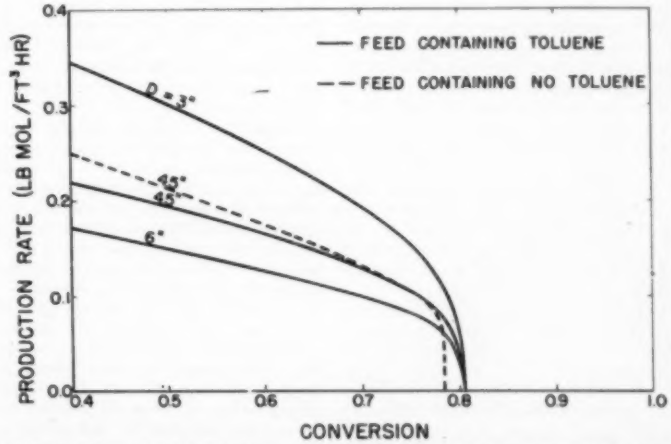

Fig. 7. Dependence of production rate on conversion with molal ratio of hydrogen to liquid equal to 5.

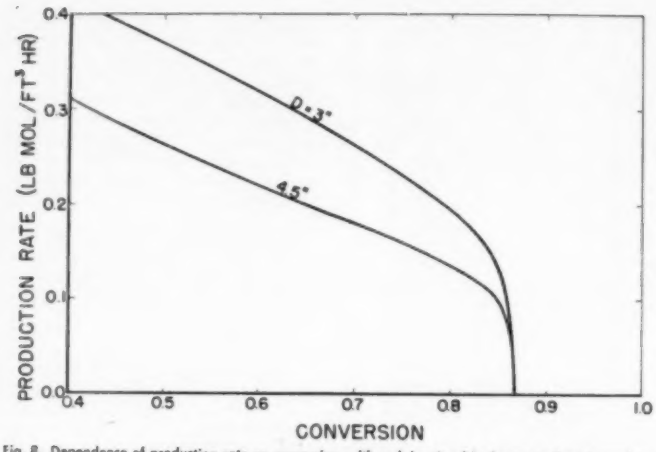

Fig. 8. Dependence of production rate on conversion, with molal ratio of hydrogen to liquid equal to 2.

that the reactor requires 38 tubes of the specified size in order to meet the conditions imposed.

After the design engineer has chosen the optimum combination of process variables and the dimensions and arrangement of the reactor, the time has come to design a pilot plant. It is nearly always necessary to operate a pilot plant before constructing a full-scale manufacturing plant for a new product or process, because there is always some uncertainty in applying empirical chemical engineering correlations, and especially because some effect of increasing the scale of the reaction may have been entirely overlooked. For example, unforeseen difficulties in the startup or shutdown procedures may come to light; or it may be found that the reaction variables selected for control are not the best, but that control of other variables or of the same variables at different points gives more stable operation. Furthermore, processes are frequently operated with recycle of some part of the effluent from the reactor, and it is usually difficult or impossible to simulate this feature acurately in bench-scale experiments. For these reasons, and others that will occur to any practising chemical engineer, one depends on the operation of a pilot plant to make evident any defects in the process design and to refine our predictions of the result of operating a commercial reactor.

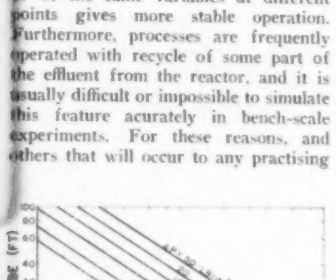

Fig. 9. Relation among flow rate, length of tube, and pressure drop.

In the case of the process discussed, the obvious choice for a pilot plant reactor is one tube of the multitube fixed-bed reactor tentatively chosen for the full-scale reactor, heated by flue gas in the same manner. By using such a reactor, with hydrogen supplied by recycle of a part of the hydrogen produced in the reaction, an experimenter can find out how he should correct his predictions of temperature profiles, and in turn correct predictions of the relation between flow rate and conversion. He discovers the effect of minor components that may build up to considerable concentrations in the recycle stream. Also some idea is obtained of the adequacy of the instrumentation provided and a good measure of the requirements for satisfactory control of the reaction conditions. Not least important, is the ability to try out small changes of the process variables to help define the optimum combination.

After the pilot plant has run long enough and with varied conditions to provide the necessary checks on the predicted performance, the design for the commercial plant can be revised by applying the corrections worked out in the pilot plant. These corrections will apply to the chemical engineering correlations of such quantities as heat-transfer coefficients and pressure drop, and also to practically unpredictable features such as the behavior of a catalyst maintained for a long time under operating conditions. Thus, the pilot plant itself furnishes the information required to increase the scale of the process on a secure basis, and the problems of scaling-up from the pilot plant are essentially the same as those involved in scaling-up the laboratory reactor to the pilot plant, with the provision that the step to the commercial scale can be made with more confidence in the light of the experience with the pilot plant.

Notation

a = activity factor, dimensionless
C_p = heat capacity of reacting fluid, B.t.u./(lb.-mole)(° F.)
D = diameter of reactor tube, ft.
F = flow rate, lb.-mole/(sq.ft.)(hr.)
ΔH = heat of reaction, B.t.u./(lb.-mole)
k = specific rate constant, lb.-mole/(cu.ft.)(hr.)(atm.)
k_e = effective thermal conductivity, B.t.u./(ft.)(hr.)(° F.)
K = pressure equilibrium constant, atm.^{3}
N_m° = mole fraction of MCH in feed to reactor, dimensionless
R = rate of reaction, lb.-mole/(cu. ft.)(hr.)
S = space velocity, lb.-mole/(cu. ft.)(hr.)
t = temperature of reacting fluid, ° F.
t_f = temperature of flue gas, ° F.
U = over-all heat-transfer coefficient B.t.u./(sq.ft.)(hr.)(° F.)
x = fractional conversion of MCH, dimensionless
z = length coordinate in reactor, ft.

Acknowledgment

The authors wish to express their appreciation to the Shell Development Co. for permission to publish this paper along with the laboratory data on toluene production. The rate data were determined in the Wilmington Laboratory of Shell Development Co.

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Discussion

George C. Gester, Jr. (California Research Corp., Richmond, Calif.): Could you estimate the amount of overdesign, for example, excess catalyst capacity, that would have been incorpo-

rated had the original estimated variables been incorporated in the commercial design without recourse to pilot plant operation?

E. Singer: The optimum design of a catalytic reactor will depend upon the cooperative effort on the part of the personnel involved. Basic data will be gathered by library research and laboratory work. These data are then processed in order to obtain a tentative optimum commercial design. This tentative design is tested then by pilot plant operation and then a final working over of the data from the basic data and pilot plant operation will yield the final design. Proper follow-through in each one of these phases is essential. I will answer by recourse to a hypothetical case, where in the design of the pilot plant we estimated an over-all heat-transfer coefficient which turned out to be in error, as shown up by the pilot plant results. I will go back to the illustrative example given in the paper, change the value of the heat-transfer coefficient, and obtain a new value for (1) the conversion using the same space velocity and reactor dimension as had been determined from the illustrative example, and (2) a new reactor design based upon the more accurate over-all heat-transfer coefficient.

Let us assume that the over-all heat-transfer coefficient was actually 10 B.t.u./(sq.ft.)(hr.)(° F.), instead of 8 which was assumed in making up the charts. This fact would show up in the pilot plant operation, and an error of this magnitude is not at all unlikely. For the same conditions as were imposed for the illustrative example given in the paper:

1. What would be the error in the calculated production rate if the reactor had the same dimensions as were given in the solution to the original problem?
2. What reactor dimensions would give the design production rate with the new heat-transfer coefficient?

In answer to question 1:

By reference to Equation (6), we note that the parameter to be considered is:

$$\psi = \frac{4UN_{m}^{\circ}}{DC_{p}a(1+DU/8k_{s})} =$$

$$\frac{4(10)(.076)}{(0.375)(14.97)(1)\left(1+\frac{0.375(10)}{8(4.4)}\right)}$$

Thus:

ψ with $D = 0.375$; $U = 10$ corresponds to ψ with $D = 0.306$ ft. (3.67 in.); $U = 8$.

We can now interpolate for $D = 3.67$ in Figure 7 which shows $f(x) = SxN_{m}^{\circ}$ (production rate) as a function of x. Since $S = 3.05$ lb.-mole/(cu.ft.)(hr.) and $N_{m}^{\circ} = 0.076$, we can find that value of x which gives $f(x)/x = 3.05(0.076)$ for $D = 3.67$ in. This turns out to be 0.69. Therefore, the production rate would be $15(0.69/0.65) = 15.9$ lb. mole/hr. which represents an error of 6%.

In answer to question 2:

From the interpolated plot for $D = 3.67$ in Figure 7 we note that at $x = 0.65$:

$$SxN_{m}^{\circ} = 0.18 \text{ lb.-mole/(cu.ft.)(hr.)}$$

$$S = \frac{0.18}{0.65(0.076)} = 3.66 \text{ lb. mole/(cu.ft.)(hr.)}$$

Since $\Delta p = 10$ lb./sq.in. and $M = 18.5$, we note from Figure 9, $Z = 21.3$ ft. Therefore, $F = 3.66(21.3) = 78.0$ lb.-mole/(sq. ft.)(hr.). Since $FxN_{m}^{\circ}(\pi D^{2}/4)n = 15$, then

$$n = \frac{15}{(78.0)(0.65)(0.076)(0.785)(0.375)^{2}} = 35.3 \text{ i.e., } 36$$

Therefore comparing the reactor dimensions for $U = 10$ with those for $U = 8$ for the same conversion, pressure drop,

U B.t.u./(sq.ft.)(hr.)(° F.)	Z ft.	$n =$ number of tubes
8	24.2	38
10	21.3	36

G. C. Gester: Did you try other types of reactors, and if so what were they, and why were they eliminated earlier in the game?

E. Singer: No, we did not use any other types. We used only the heat-exchanger type. Other types are in operation at the present time, notably the adiabatic type. Our decision to use the heat-exchanger type was based on economic considerations evolving from the type of analysis illustrated in the paper.

G. C. Gester: On the basis of your rate equations for dehydrogenation of methylcyclohexane, you indicated the reverse reaction was of the fourth order. In my experience that is quite a rarity, and generally, such reactions occur as a succession of lower-order reactions which series of lower-order reactions give an apparent fourth-order reaction. Have you had a chance to fit the data to successive curves of lower-order reactions?

E. Singer: Probably the reaction is more complicated than is indicated by the analytic expressions. We have indicated that the forward reaction is first order in methylcyclohexane. If it is true that forward reaction is first order over the entire concentration range, it must follow from thermodynamic considerations that the reverse reaction is fourth order. The forward reaction is probably more complicated than is pictured, since it involves the adsorption of methylcyclohexane on the catalyst, its subsequent reaction on the surface of the catalyst, and then the desorption of the reaction products. If we do grant, however, that the forward reaction is truly first order in methylcyclohexane, the reverse reaction must then be fourth order. In this case, for a heterogeneous catalytic reaction, the probability of the meeting of four reactant molecules on the surface is quite a bit higher than in the gas phase, since we are dealing with an adsorbed phase on the surface of the catalyst. The most decisive answer to your question is that over the range of concentration and conditions of interest, the analytical expression shown in the body of the paper describes the rate of reaction.

D. M. Hurt (Du Pont Co., Wilmington, Del.): Though we are in agreement with the methods detailed by the authors, it might be worth mentioning that there are some alternate techniques which might be simpler for some limited objectives.

First, if we recalled the rate equations given in the classical form as the difference between the forward and backward reactions, by inspection only, and not having tried it, it would appear that in the case used, since it was found necessary or desirable to employ an excess of hydrogen for use in the catalysts, a simple, apparent first-order equation might be equally good. This, of course, is only an expediency. The advantage would be that it would eliminate a large amount of algebraic work. Next, the authors method of approach was, I might say, a rather general one in that they prepared a large number of angles and curves on which any specific solution could readily be developed. This is perhaps a matter of choice and depends, in the long run, entirely on the number of calculations one might want to make. The ultimate, of course, would be to prepare one or two direct analytical solutions in which you join the data for the specific points you are evaluating.

We agree entirely in principle, at least, with the authors as to the desirability of first designing the commercial scale equipment before building the pilot plant.

In the first place—why build a pilot plant? There are, of course, in some cases, obvious and sufficient reasons, such as desirability of producing material for evaluation, or materials for pilot testing. Some of the physical operations subsequently involved but leaving those to one side, the main objective in this case, at least, is to confirm the operation of the commercial design. Obviously, what commercial design are we talking about? We are talking about the optimum or economic optimum commercial design and if the pilot plant is to fulfill its function, it should be a prototype of the ultimate plant and not something that is merely intermediate in scale.

There is one point that comes up in this specific type of equipment—the multitubular reactor which points out a

lack of knowledge in an allied field, and is considered well developed—that is, heat transfer. A possible point of trouble in this type of equipment and which is more serious in the exothermic reaction than in the endothermic, is the inequality of temperature distribution or the inequality of gas flow on the shell side of the exchanger.

You are all familiar with the work on heat transfer. We have methods for calculating coefficients on cross flow and when we get through we apply a correction factor of 0.6 to take care of the internal by-passing on the shell side.

This can have a serious effect on multitube reactors. It points out a lack of knowledge for which we do not yet have the answers. We have, in some cases, gone to the extreme of running excessively large amounts of cooling fluid at low-temperature gradients to obviate this difficulty. Finally, there is one point on which the author might wish to make some comment, and that is the extent to which basic data are used. The objective of obtaining complete basic design information or rate data on any process is, of course, to permit design of not only any size equipment but also any type of equipment—in this case tubular reactors—the adiabatic type of single or multiple-bed or some of the moving-bed types. From the data given, it is possible to calculate the performance of any of these types.

Normally, we depend on judgment—we might call it intuition or experience—to reduce the number of examples we wish to evaluate. There are some factors which enter into this judgment: if say, the judgment of the engineering department happened to coincide with the judgment of the operating department on who will run the plant, the preferred type, it would probably not be necessary to evaluate the alternates. Have you run into this situation and would you care to comment on it?

E. Singer: Mr. Hurt's first question was pertaining to the technique used in the calculation, and he suggested that the over-all reaction might be approximated by an apparent first-order reaction. The complete expression, even though it might appear complicated, is necessary, since without the backward reaction there would be no equilibrium as far as chemical reaction is concerned. The person doing the calculations will note that at the beginning of the bed, where conversion is practically zero the reverse reaction is negligible, thus simplifying his calculations considerably. Towards the end of the bed where the conversion is not inconsiderable, the reverse reaction does assume important proportions.

Mr. Hurt suggested that the number of calculations and working charts would be decreased considerably if one could get an analytical expression for the average conversion, let us say, in terms of the length of the bed, the temperature of the flue gas, etc. I agree; however, the expression would be quite complicated if obtainable at all. We tried to get such an analytical solution and failed, so we were obliged to go into these working charts in order to find an adequate solution.

I would like to underline Mr. Hurt's comment on our ignorance in the field of heat transfer in catalytic reactors, but instead of talking about the shell side of the reactor, let me talk about the inside of the tube.

As you will note in our analysis we broke up the heat-transfer resistance into two parts: the resistance of the bed itself, and the film at the wall. We used this procedure in our analysis without any good justification. The reason for this procedure was that the resistance at the outside of the tube was the major one, so that our ignorance of this point did not show up. There are no conclusive data in the literature at present, and I do not think there is any university work now under way on a good analysis of this question.

Mr. Hurt then mentioned that considerable data had to be assembled in order that one might make an economic balance on all the possible types of reactors. Our paper presented an analysis for only one type. It is true that much fundamental data must be assembled in order to analyze all possible types of reactors. On the other hand, in many instances, a clear-cut decision may be made after only a cursory examination of the reaction. It may be immediately seen, in some instances, that the economics of one reactor type are much more favorable than others, or that one type is much more easily controlled or may be capable of more flexibility in throughput than any of the other possible types. Each reaction, of course, must be treated individually.

Louis S. Kassel (Universal Oil Products Co., Chicago, Ill.): This paper treats a problem which is simpler than would usually be encountered. This may be seen by considering an imaginary process identical with that discussed, except that all temperatures are 300° F. lower. In this imaginary process it will be fair to assume that the flue gas temperatures are less than the 1200° F. limit imposed by mechanical engineering considerations. In the actual process production rate is a function of tube diameter and the optimum diameter is determined by an economic study of this function. In the imaginary process production rate may be arbitrarily selected. This production rate may be achieved for a range of tube diameters. Conversion efficiency, however, is a function of tube diameter and the optimum diameter is determined by an economic study of this function. The conversion efficiency is calculable in principle from differential reactor data. In practice it is difficult to obtain adequate differential reactor data and tedious to analyze it. These are the difficulties which the authors avoided by stating that no large amount of the side reactions can be tolerated. Without the 1200° F. flue gas temperature limitation, even minor side reactions play a major role in the design problem.

E. Singer: The actual design that was used is far from the theoretical conditions that Mr. Kassell outlined, but it is extremely illuminating to point up difficulties that might occur in a more complicated case. We have chosen this simple example for purposes of illustration of method. The guiding principles that have been outlined are applicable to any case. For instance, if the reaction is so complicated that an analytical expression for the rate of reaction is not obtainable, then the tabulated or graphical data obtained from differential reactor runs may be used to establish R for use in the expressions given.

Roland Voorhees (Carbide and Carbon Chemicals Co., S. Charleston, W. Va.): I was interested in the apparent fourth-order mechanism for the reverse reaction. I suspect that probably the true mechanism of this reaction is first order, and that the mechanism appears to be fourth order just because of the mathematics involved.

The driving force is really the difference between the actual concentration and the concentration at equilibrium. If the reaction appears to be fourth order, it may be because of the nature of the equilibrium constant. In other words, if you express the rate of reaction as rate of change, not of concentration, but of concentration minus concentration at equilibrium, with respect to time, you would probably find that both the forward and the reverse reactions, expressed in this manner, were first order. If the concentration at equilibrium is translated into terms of the equilibrium constant, the mathematics involved may have the effect of making the reverse reaction appear to be fourth order. Have you or has anybody ever tried to hydrogenate toluene, starting with pure toluene; will the data obtained in this case plot as a fourth- or first-order reaction?

E. Singer: I refer you to the discussion of Mr. Gester's questions for the answer to your first question. No, we have not made any runs on the hydrogenation of toluene.

(Presented at White Sulphur Springs (W.Va.) Meeting.)

THE PRANDTL NUMBER FOR STEAM

FRANK L. RUBIN

The Lummus Company, New York, New York

THE Prandtl number $c_p\mu/k$, which is dimensionless, is a basic function in the determination of heat transfer by convection.

Many assumptions have been made in regard to the numerical value of the Prandtl number, e.g., ". . . . the term $c_p\mu/k$ is assumed independent of pressure" (*8*) and ". . . . $c_p\mu/k$ equals 0.78 for steam . . . ." (*9*) and "The Prandtl number . . . . is nearly the same for all gases over a wide range of temperature" (*1*). Various equations have been developed for heat transfer of gases based on these and similar premises.

The Prandtl number for steam is plotted as a function of temperature and is tabulated here in Table 1. At low pressure the Prandtl number is relatively constant at a dimensionless value of 1.0. At intermediate and high pressures it varies with both temperature and pressure. This variation extends from 1.0 to 5.3. Furthermore it is likely that one "will find other substances are similar in this relationship to the course for steam (*4*)." It now becomes necessary to re-examine those relationships, which have been based upon certain assumptions in regard to the Prandtl number.

The values of the Prandtl number have been determined from data in the twenty-first printing (June 1950) of "Thermodynamic Properties of Steam" (*3*). Recently published papers present new data for the thermal conductivity of steam (*5, 6*) and new relationships for the viscosity of steam (*2, 6, 7*). This material has been incorporated into Table 6 for Viscosity and Table 7 for Thermal Conductivity in this latest steam table (*7*). The specific heat can readily be calculated from the enthalpy values of Table 3 (*3*) for superheated vapor.

Only at those pressures and temperatures for which the thermal conductivity and viscosity are tabulated has the Prandtl number been calculated. The specific heat was determined by using the enthalpy data at the specified pressure and for the temperatures in the table immediately above and below that specified. (The specific heats at zero and one pound per square inch absolute were considered equal. The specific heat at the saturation temperature was calculated from the enthalpy at saturation and at the lowest, superheated temperature at that pressure.)

μ = viscosity, (lb. force)/(sec.)(sq.ft.)

c_p = specific heat, B.t.u./(lb. mass)(° F.)

k = thermal conductivity, B.t.u./(hr.)(sq.ft.)(° F./ft.)

g_c = conversion constant, (lb. mass)(ft.)/(sec.)^2(lb. force)

3600 = sec./hr.

The units of viscosity and thermal conductivity are those of the steam tables (*3*). The units of specific heat are those directly calculated from the enthalpy data. The dimensionless Prandtl number is given by

$$Pr = \frac{c_p\mu 3600 g_c}{k}$$

It is estimated that the Prandtl number values calculated by this method are accurate within about two per cent.

Acknowledgment

The author extends his thanks to Frederick G. Keyes for his research which made this paper possible.

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TABLE 1.—PRANDTL NUMBER ($C_p\mu/k$ Dimensionless)

Abs. Press. lb./sq.in.	Sat. Vapor	Temperature—° F. 200°	400°	600°	800°	1000°
		1.063	1.022	1.003	1.002	.995
500	1.90	.....	.....	1.40	1.19	1.12
1000	2.67	.....	.....	2.06	1.44	1.27
1500	3.81	.....	.....	3.53	1.73	1.42
2000	5.27	.....	.....	.....	2.09	1.60

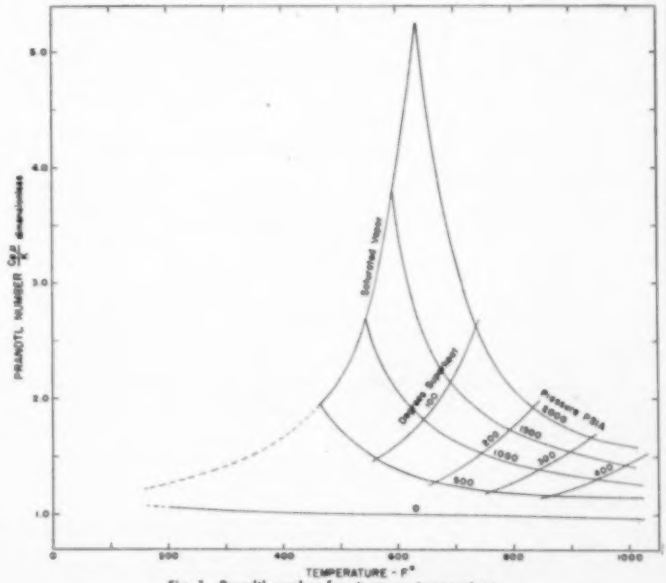

Fig. 1. Prandtl number for steam vs. temperature.

CAST ALLOY REFERENCE SHEET

N. S. MOTT, Chief Chemist Metallurgist
The Cooper Alloy Foundry Co., Hillside, N. J.

ALLOY: CORROSION-RESISTING LOW CARBON 18% CHROMIUM 8% NICKEL ALLOY.

APPLICATIONS AND REMARKS: The well-known 18-8 type is a general purpose corrosion-resistant alloy for applications in chemical, pharmaceutical, textile, and food-processing industries, and in oil refineries. Although the base cost for castings of this material is about 58 cents/lb. above that for carbon steel, the added long life and guarantee of product purity will quite often justify the high initial cost. Carbon is held low in this alloy for maximum corrosion resistance. It is cast into valve parts, pump parts, fittings, and a wide variety of cast shapes for the above-mentioned industries.

DESIGNATIONS: ACI CF-8; AISI 304; SAE 60304; ASTM A296-49T, Grade CF-8.

CHEMICAL COMPOSITION RANGE: C < .08%; Cr 18-21%; Ni 8-11%; Si < 2%.

MACHINABILITY: 18-8S is tough, ductile, and is characterized by its work hardening and high frictional properties. Best results for turning are obtained by using a slow heavy cut. To avoid work-hardening the surface, care should be taken not to allow the tool to ride or glaze the work. Sharp tools with a large rake are desirable, and chip-curler tools are recommended.

HEAT TREATMENT: If carbides are not put into solution intergranular corrosion may result. To avoid this all castings are water-quenced from 1950-2050° F. before being put into corrosive service. The time at temperature is from 1 to 3 hours, depending upon metal thickness.

WELDABILITY: 18-8S may be readily welded by the usual methods using a type 304 filler rod or preferably type 308 (20-10S). In the welded condition, however, it is subject to intergranular corrosion or weld decay in areas adjacent to the weld deposit. To avoid this, water - quenching of the work from 1950-2050°F. is necessary.

MECHANICAL AND PHYSICAL PROPERTIES:

Tensile Strength, 1000 lb./sq.in.	75	Melting Point	2550-2650 F.
Yield Point, 1000 lb./sq.in.	45	Specific Heat (B.t.u./(lb.)/° F.) 32-212° F.	0.12
Elongation, %	55	Thermal Expansion (× 10⁻⁶in./(in.)/° F.) 32-212° F.	9.6
Reduction in Area, %	55	Thermal Conductivity (B.t.u./(hr.)(sq.ft.)(° F./in.)) 32-212° F.	113
Brinell Hardness	140	Electrical Resistance (Ohms./cir. mil. ft.) @ 32° F.	420
Charpy Impact (Std. Keyhole ft.-lbs.)	75		
Mod. of Elasticity (× 10⁶ lb./sq.in.)	28		
Specific Gravity	7.92		

CORROSION RESISTANCE

ACIDS	
Acetic 5%, 70 F	E
Acetic 5%, boiling	E
Acetic 80%, 70 F	E
Acetic 80%, boiling	F
Acetic Glacial, 70 F	E
Acetic Glacial, boiling	F
Benzoic 5%, 70 F	E
Boric 5%, 176 F	E
Chromic 10%, 70 F	E
Chromic 10%, boiling	F
Chromic 50%, boiling	P
Citric 5%, 70 F	E
Citric 25%, boiling	G
Citric 50%, boiling	F
Formic 5%, 70 F	G
Hydrochloric 1%, 70 F	P
Hydrochloric 1%, boiling	N
Hydrochloric, 5%, 70 F	N
Hydrochloric 5%, boiling	N
Hydrochloric 25%, 70 F	N
Hydrochloric 25%, 176 F	N
Hydrofluoric 48%, 70 F	P
Hydrofluoric 48%, 176 F	N
Lactic 5%, 70 F	E
Malic, all temps.	E
Nitric all concs., 70 F	E
Nitric 65%, boiling	E
Oleic all concs., all temps.	E†
Oxalic 5%, boiling	E
Phosphoric 10%, 70 F	E
Phosphoric 85%, 70 F	N
Phosphoric 85%, boiling	N
Stearic concentrated to 200 F	E
Sulfuric 2%, 70 F	G
Sulfuric 2%, 176 F	N
Sulfuric 2%, boiling	N
Sulfuric 5%, 70 F	F
Sulfuric 5%, 176 F	P
Sulfuric 5%, boiling	N
Sulfuric 10%, 70 F	P
Sulfuric 10%, 176 F	N
Sulfuric 10%, boiling	N
Sulfuric 78% (60° Be), 176 F	N
Sulfuric 93% (66° Be), 70 F	N
Sulfuric 93%, 300 F	G
Oleum, 70 F	G
Mixed Acids 57% H_2SO_4, 28% HNO_3, 176 F	G
ALKALIES	
Ammonium Hydroxide, all concs.	E
Calcium Hydroxide Sat., 70 F	E
Calcium Hydroxide Sat., boiling	F
Sodium or Potassium, Hydroxide, all concs., 70 F	E
Sodium or Potassium Hydroxide < 20%, boiling	E
Sodium or Potassium Hydroxide 30% boiling	F
Sodium or Potassium Hydroxide, Molten, 600 F	N
NEUTRAL AND ALKALINE SALTS	
Barium Sulfide, 70 F	G
Calcium Chloride 5%, 70 F	G†
Calcium Sulfate Sat., 70 F	E
Magnesium Chloride 5%, 70 F	G†
Magnesium Sulfate 5%, 70 F	E
Sodium Carbonate, all concs., 70 F	E
Sodium Chloride 5%, 70 F	G†
Sodium Sulfate 5%, 70 F	E
Sodium Sulfide 5%, 70 F	E
Sodium Sulfite 5%, 70 F	G
ACID SALTS	
Alum 10%, boiling	G†
Aluminum Sulfate 10%, 70 F	G†
Ammonium Chloride 5%, 70 F	G†
Ammonium Sulfate 10%, 70 F	G
Ammonium Sulfate 10%, boiling	F†
Ammonium Nitrate, all concs., 70 F	E
Stannous Chloride 5%, 70 F	F†
Zinc Chloride 5%, boiling	P†
OXIDIZING ALKALINE SALTS	
Calcium Hypochlorite 2%, 70 F	F†
Sodium Hypochlorite 5%, 70 F	F†
Sodium Peroxide	E
OXIDIZING ACID SALTS	
Ammonium Persulfate 5%, 70 F	E
Cupric Chloride 1%, 70 F	P†
Cupric Sulfate 10%, 70 F	E
Ferric Chloride 10%, 70 F	P†
Ferric Sulfate, boiling	G
Mercuric Chloride 2%, 70 F	P†
Stannic Chloride 5%, 70 F	P†
WET AND DRY GASES	
Chlorine Gas Dry, 70 F	G
Chlorine Gas Wet, 70 F	N
Sulfur Dioxide Dry, 575 F	E
Sulfur Dioxide Wet, 70 F	G
Sulfur Dioxide Solution, 70 F	F
Sulfur Dioxide Spray, 70 F	P†
Hydrogen Sulfide Dry	E
Hyrdogen Sulfide Wet	G
ORGANIC MATERIALS	
Acetone, 70 F	E
Acid Sludge (50% H_2SO_4), 200 F	N
Alcohol—Methyl and Ethyl	E
Aniline Hydrochloride, 70 F	N
Benzol, 176 F	E
Carbon Tetrachloride	E
Chloroform	E
Ethyl Acetate, 70 F	E
Formaldehyde, 70 F	E
Phenol 5%, boiling	E
Refinery Crudes	F
Trichlorethylene, boiling	E
PAPER MILL APPLICATIONS	
Kraft Liquor	E
Black Liquor	E
Green Liquor	E
White Liquor	E
Sulfite Liquor, 176 F	G
Chloride Bleach	F†
Paper Makers Alum	G
PHOTOGRAPHIC INDUSTRY	
Humid Atmospheres	E
Cellulose Acetate	G
Acetic Anhydride	G
Acetic Acid + 1% H_2SO_4	P
Developers	E
Solutions Containing SO_2	F
Silver Nitrate, 70 F	E
FERTILIZER MANUFACTURING	
$H_3PO_4 + H_2SO_4 + HF$	P
PICKLING OPERATIONS	
H_2SO_4 + Dichromate, 176 F	G
H_2SO_4 + Hcl, 176 F	N
CORROSIVE WATERS	
Acid Mine Water	G†
Abrasive Acid Mine Water	P
Sea Water	G†
Brackish Water	G†
FOOD & ASSOCIATED PRODUCTS	
Brines	G†
Edible Oils	E
Fats	E
Fatty Acid Distillation	E
Fruit Juices	E
Ketchup	G†
Milk Pasteurizing	E
Vinegar and Salt, 70 F	G†

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F—Fair resistance. 0.005-0.020 in. of penetration per year. Satisfactory service under specific conditions. Light to moderate attack.
P—Poor resistance. 0.020-0.120 in. of penetration per year. Satisfactory for temporary service only.
N—No resistance. 0.120 min. in. of penetration per year. Rate of attack too great for any use.
†—Subject to pitting type corrosion.

No. 4
(FOURTH OF A SERIES OF TEN)

ATLANTIC ANNUAL ALL AWARDS AT CITY CONVENTION COME CONFERENCES CHALFONTE-HADDON HALL

Article by L. C. Knox

IN all its 43 years of annual meeting-going, the American Institute of Chemical Engineers has made it to Atlantic City only once by design (1931), and once by default (1946) when a last-minute hotel strike sent a scheduled Philadelphia meeting off to the shore. But this neglect of South Jersey's convention town *par excellence* ends the month after next when A.I.Ch.E. members and many others gather at the Chalfonte-Haddon Hall Hotel on Dec. 2 through Dec. 5 for their Forty-fourth Annual Meeting. There some 1500 to 2000 of them will transact business, hear the Third Institute Lecture, attend five symposia and a panel discussion, present their deserving fellows with annual awards, and be entertained (the ladies too, of course) in a well-planned round of sociability. A quick glance at the following program reveals a crowded four-day get-together.

Thomas E. Murray, speaker at Awards Banquet.

ARE UNIT OPERATIONS OBSOLETE?

The question in the title is far from idle rhetoric and the demise of this important concept in chemical enineering might be nearer than we think. A full discussion on the changing concepts in our profession will be given at the Annual Meeting at Atlantic City, Dec. 2-5, 1951, at a panel, "Whither Chemical Engineering Science," to be headed by George Granger Brown. This is one of the questions that will come up and in preliminary correspondence on the matter one member has said, "While the unit operations concept has been essential and invaluable to the current development of chemical engineering, it is no longer needed, and if arbitrarily continued may retard rather than assist future development of chemical engineering science.

Basic concepts such as this will be discussed at the Annual Meeting. Plan to attend.

Registration

Sunday. Registration, first as always, goes on from 1:00 to 7:00 P.M. on Sunday for all the early arrivals and all day on each subsequent day of the meeting in the English Lounge of Haddon Hall. Members can speed up the necessary mechanics immensely by preregistering. Simply filling out and mailing the post card enclosed in each program announcement will eliminate practically all the queueing-up and name-and-address repeating usually associated with the process.

The Third Institute Lecture

Monday. This year, the recently established Institute Lecture will be given by Thomas B. Drew of Columbia. The lecture is scheduled for Monday morning, immediately before the business meeting. Professor Drew's topic will be "Diffusion." His association with researches in these fields has been a long-standing one. His address will continue the tradition, already established by W. H. McAdams and O. A. Hougen in the past two years, of presenting annually concise, authoritative résumés in the several departments of chemical engineering.

The Awards Banquet

Tuesday. At Tuesday evening's Awards Banquet, aside from watching the formal installation of the new officers, members will hear this year's distinguished guest speaker, Atomic Energy Commissioner Thomas B. Murray, and applaud the winners of the three regular Institute awards: the Student Contest Problem prizes, the William H. Walker Award, and the Junior Award; and a fourth, the Professional Progress Award, granted by the Celanese Corporation of America, but administered by the Institute.

(Continued on next page)

Well-known Absecon Lighthouse.

The Student Contest Problem prizes, of which the A. McLaren White Award is the first, are the Institute's oldest awards, founded 20 years ago. Last year's first prize winner was Robert P. Bannon, of the University of Illinois.

The William H. Walker Award and the Junior Award have a similar purpose in that both are awarded for outstanding papers contributed to the Institute. The second award is given only to Junior members. In 1950, the winners of the two prizes were B. F. Dodge of Yale and F. M. Tiller of Vanderbilt University, respectively.

An outstanding contribution to chemical engineering leading to the betterment of human circumstances is honored by the Professional Progress Award, a grant of $1,000 and a certificate, given by the Celanese Corporation of America. Prof. E. R. Gilliland last year received this award. This year, the medalist, instead of addressing the Awards Banquet, will speak to the Institute on Tuesday, before the morning session.

The speaker at the Awards Banquet will be Thomas E. Murray, member of the Atomic Energy Commission. Mr. Murray is a mechanical engineer who became an electrical engineer and inventor. For 35 years he was a full-time business executive and part-time public servant who handled a number of special labor relations jobs for the Government, but now he is devoting all of his time to the Atomic Energy Commission since his appointment by President Truman in 1949.

The Technical Program

This year's technical program, besides a full complement of general papers, comprises five symposia and an informal panel discussion.

The informal discussion, on Sunday afternoon, poses the question, "Whither Chemical Engineering Science?" Moderator George Granger Brown and a panel of five from the universities and industry will try to answer it as they discuss the effect of latter-day findings on the division of chemical engineering into the "classical" unit operations.

The Symposium of Heat Transfer, with D. L. Katz and A. S. Foust as cochairmen, will be preprinted. Copies will be available before and during the meeting. (See Secretary's Report for titles of papers.)

The subjects of L. P. Scoville's session on Effective Speaking are required reading for all engineers, chemical species not excepted. One paper concerns rapid reading and the other effective oral presentation.

Horseback riding on the beach replaces seasonal swimming.

Five papers are scheduled for the Symposium on Industrial Waste Disposal presided over by R. D. Hoak of Mellon Institute. This is a continuation of last year's session at Columbus.

The part the chemical engineer plays in the marketing organization is the general topic of the five papers to be presented in a Symposium on Opportunities in Sales for Chemical Engineers, with F. J. Curtis of Monsanto as chairman.

Another Symposium on Chemical Engineering Fundamentals will have James Coull of Pittsburgh and W. Edmister of Carnegie Tech. as cochairmen. Four papers will be presented at this session.

The Lighter Side

Sunday night's informal open-house is the official ice-breaker for all the early comers, including the ladies. A small group of entertainers for this one is promised, and the standard assortment of refreshments, in the Vernon Room of Haddon Hall after 8:30.

Monday night is Casino Night, when everybody gets a chance to become a large-scale plunger quite painlessly. The Kefauver Committee will never look for the $10,000 you'll get to blow in on blackjack, craps, birdcage, roulette, horse racing and poker. Prizes if you're lucky in Haddon Hall on the Tower Floor at 9:00.

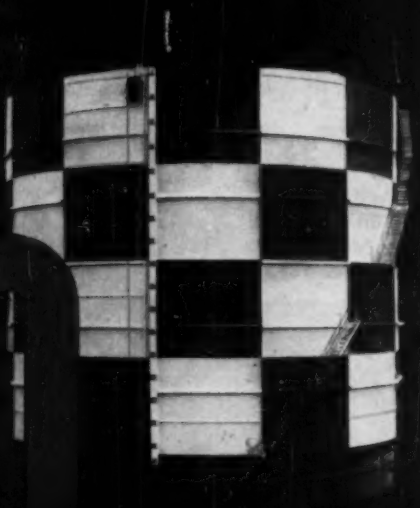
Especially for the Ladies

The Sunday, Monday and Tuesday night affairs are meant for the members and the ladies, but as usual, the ladies have another program designed especially for them. There's coffee in the morning, and tea in the afternoon of each day, and at least one special diversion every day.

On Monday afternoon, for example, there's a demonstration of what millin-

(*Continued on page* 59)

A view of the famous boardwalk.



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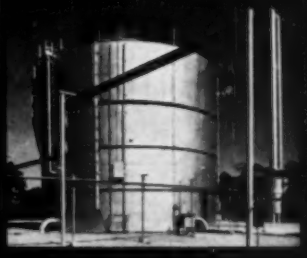

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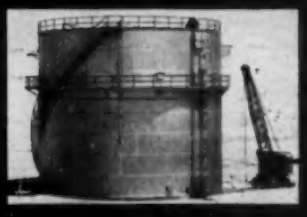

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ROCHESTER REPORT

Registration was heavy. The local committee expected around 350, the actual total came to 635.

Publicity for the highly successful Rochester meeting was handled by chairman E. K. McLean, left, of Distillation Products ind., W. F. Kaiser, Eastman Kodak Co., and R. J. Gustafson, Rochester Gas & Electric Corp.

Some members of the hard-working Registration Committee; right to left, Earl W. Costich, committee chairman, Mixing Equipment Co.; E. G. Waltz and Ed. Garretson of Eastman Kodak Co. Standing: Tom Galley, a member of the Hospitality Committee who manned the information desk, also of Eastman Kodak.

THE home of photography in the United States was visited during the week of September 16 by the nation's chemical engineers who turned out in such force that Rochester now ranks as the city which attracted the fifth largest regional gathering of the Institute. The program, kept small as part of the Institute's policy of making the meetings attractive to the local and visiting engineers, drew an attendance of 635.

The meeting applauded D. E. Pierce for his competence as technical program committee chairman.

The pattern of Institute meetings of an earlier day was followed, inasmuch as no technical meetings were scheduled for the first two afternoons of the week, allowing this time to be used for plant trips. Attendance was high and interest in the trips never slackened.

The program had many unusual highlights—the first, the symposium on, "Maintenance of Chemical Plants," with D. E. Pierce, General Aniline & Film Corp., Chairman of the Rochester technical program, presiding. This proved to be an attraction for hundreds of plant maintenance engineers. Discussion was lively and many practical problems that come up in every-day maintenance were discussed both during the meeting and following the presentation of the papers.

Another feature which brought out extensive discussion and kept the engineers talking for hours after the meeting finished, was an evening round-table discussion on, "The Engineering Manpower Shortage and What We Can Do About It."

Other highlights of the Rochester

(*Continued on page 22*)

Bus line-up for the Eastman Kodak plant trip. Committeemen collected all lighters, matches and cameras, kept them in the bag for the duration.

A table at the Sunday night get-together included, center left—around table, R. F. Elmer, Pfaudler Co.; H. W. Fry, Pittsburgh Coke and Chemical Co.; R. E. Smith, Pfaudler Co.; I. Siller, Pfaudler Co., and Mrs. Siller, A. J. Bruno, Pittsburgh Coke and Chemical Co.; Ted Dunford, H. Raut, H. J. Greene, E. W. Neben, J. M. Culotta, all of the Pfaudler Co.

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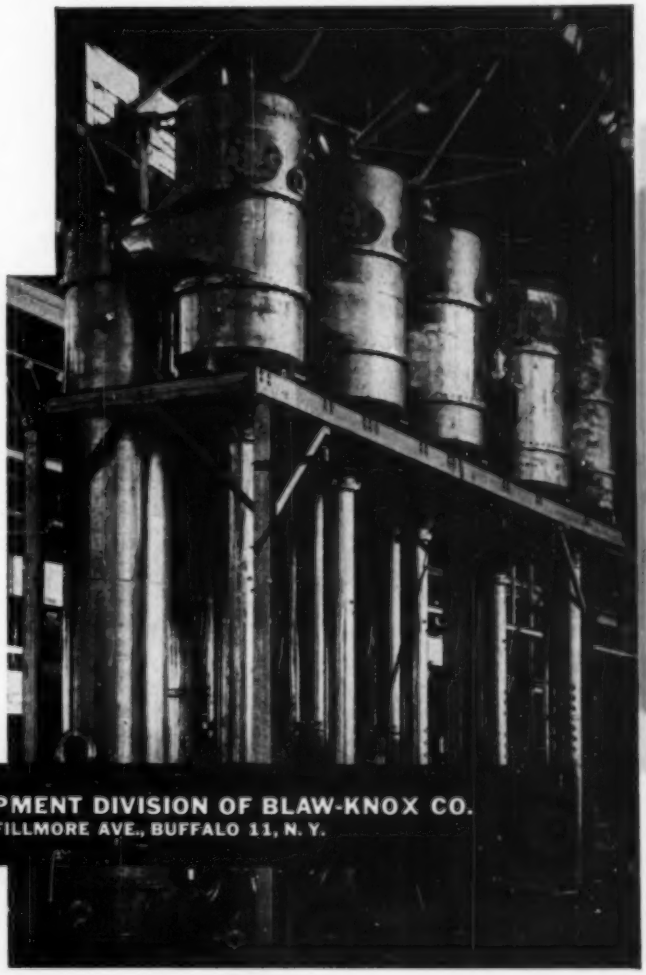

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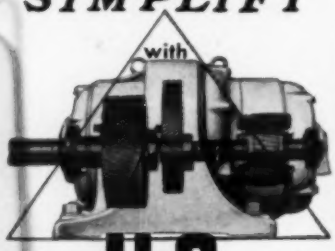

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ROCHESTER MEETING

(*Continued from page 20*)

The ladies relaxed with tea and a style show, after a busy schedule of sightseeing and plant trips.

meeting were the symposium on, "Industrial Application of Photography," a round-table discussion on, "Chemical Engineering Education," under the direction of W. E. Gift, and an evening inspection tour of the George Eastman House, a shrine to the early pioneer in photography and now a museum of photographic history.

The winner of the Institute prize for the best presented paper at the Rochester meeting was Richard D. Boutros, chief engineer, Mixing Equipment Co., Inc., Rochester, N. Y., for his delivery of the paper, "Installation, Operation, and Maintenance of Fluid Mixing Equipment." The judging committee of eight was headed by C. M. Sliepcevich, of the University of Michigan.

In the manpower symposium Carey H. Brown, chairman of the Engineering Manpower Commission of Engineers' Joint Council, led the discussion. His talk was a résumé of the present shortage of engineers and the general manpower situation. He said that in 1951 and for 1952, the need for engineering graduates has been estimated to be 95,000 men to fill places in industry and in the armed services. "Poorly balanced against this need," he said, "is the supply of 38,000 graduates in 1951 and 23,000 for 1952. The estimated needs taper off gradually to about 55,000 a year by 1960 with the estimate of prob-

Tape-recorded rehearsal for radio broadcast by leading Institute members. Left to right, William Burt, Vice-President A.I.Ch.E., A. K. Ackoff, general chairman, Rochester meeting, T. H. Chilton, President, A.I.Ch.E., and S. L. Tyler, Executive Secretary, A.I.Ch.E.

Chemical engineering luncheon, Hotel Sheraton, center left—around table with Edwin G. Willcox and J. W. Hemphill of Johns-Manville, C. R. Nash, A. L. Adams, G. R. Clark and G. H. Stark, all of Eastman Kodak Co.

A cocktail party on the Starlight Roof preceded the Tuesday evening banquet.

able graduates leveling off at about 18,000 from 1953 on," he continued. It was his feeling that this situation is critical because engineers are needed to supply superior weapons to our armed services in order to offset the numerical strength of our enemies, and to maintain a sound civilian economy sturdy enough to support our military program.

The second speaker, Dwight W. Durham of Du Pont, made the point that while industry is aware of the engineering manpower shortage problem, each company may be relying on meeting its own needs, and there has not begun as yet any unified movement to recognize the problem on an over-all level and act to alleviate the shortage on a national level. He also mentioned the shortage should be publicized as widely as possible so that the public, the government, and industry, will start to take effective action. He suggested closer contact between industry and high schools in order to obtain a fair share of the students for engineering training, and then suggested that in the immediate future engineers be used more effectively by industry.

A particularly challenging talk in the manpower round table was given by Dean G. G. Brown of the University of Michigan. Early in his remarks he said that engineers now graduating, "are getting almost as much as an inexperienced plumber."

O. E. Dwyer, Univ. of Rochester, Richard D. Boutros, Mixing Equipment Co.

Brown also said it was industry's job to increase the supply of engineers and challenged industry by asking, "Who wants these engineers, anyway? How badly do you want them? Do you want them bad enough to keep them out of the armed forces? Do you want them bad enough to back strongly the recommendation of the advisory committee? Do you want the extra engineers badly enough to go out to the high schools and beat the bushes and tell the fact that one of the best ways to become an administrator of industry is through the route of an engineering education? Do you want to go hard enough to say that one of the poorest ways to become a business administration? Are you willing to go to

(Continued on page 39)

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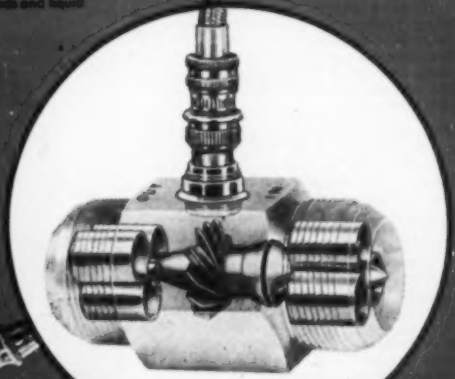

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PROGRESS AWARD WINNER, INSTITUTE LECTURER ANNOUNCED

KIRKBRIDE WINS PROGRESS AWARD

Chalmer Gatlin Kirkbride, vice-president and director of Houdry Process Corp., is the 1951 winner of the Professional Progress Award in Chemical Engineering, sponsored by the Celanese Corporation of America and administered

C. G. Kirkbride

by the A.I.Ch.E. The award which carries with it $1000 in cash, has been given to Mr. Kirkbride, who is at present in charge of research and development for the Houdry Process Corporation,

> "For his many contributions and gifted leadership in Chemical Engineering education and practice, for his scientific developments underlying petroleum technology, and for his tireless and unselfish service to the Profession."

The presentation will be made at the Awards Banquet at the annual meeting of the A.I.Ch.E at Atlantic City in December. The Award address will be made before a technical session of the Institute earlier the same day.

The Institute Awards Committee, headed by Allen P. Colburn, provost of the University of Delaware, made the selection. The award is made each year to a technical engineer under 45 years of age, "to recognize outstanding progress in the field of chemical engineering for the betterment of human relations and circumstances."

Kirkbride is a graduate in chemical engineering with a degree of M.S. in chemical engineering from the University of Michigan, and has been with the Standard Oil Company of Indiana, assistant director of research with Pan American Refining Corp., chief of chemical engineering development, Magnolia Petroleum Co., and professor of chemical engineering at A. & M. College of Texas.

He is the author of several papers on heat transfer, multicomponent fractionation and phase equilibria, and holds patents on desalting crude petroleum and high pressure absorption of hydrocarbons. He is the author of the book "Chemical Engineering Fundamentals," published in 1947.

In addition to his membership in the A.I.Ch.E., he is a member of the American Petroleum Institute, American Society for Engineering Education, A.C.S., and The Chemists' Club.

He has been chairman of the Membership Committee, Public Relations Committee, and Program Committee for the A.I.Ch.E., as well as a Director (1946-48), and is a present Director with a term expiring in 1952.

Kirkbride was a nonparticipating scientist invited to attend the atom bomb test at Bikini, and members will remember his series of articles on this expedition published in the "A.I.Ch.E. Bulletin" in 1946.

He has been an active member of E.C.P.D., E.J.C., and is a member of Tau Beta Pi, Phi Lambda Upsilon ,and Alpha Chi Sigma.

INSTITUTE LECTURE BY DREW

Thomas Bradford Drew, head of the department of chemical engineering at Columbia University, has been chosen to dealiver the third Institute Lecture at the annual meeting of the A.I.Ch.E. in December. His talk will be on, "Diffusion," and will be given prior to the annual business session on Monday morning.

T. B. Drew

Professor Drew has long been associated with the chemical engineering phenomena of diffusion, and the Institute Lecture is recognition of his authority in this field. The lecture will form the basis for a volume on diffusion in the Chemical Engineering Progress Monograph Series. It will be published next year.

The Institute Lecture was begun in 1949 when Prof. W. H. McAdams delivered a lecture on "Heat Transfer," which was followed by last year's lecture by Olaf A. Hougen on "Reaction Kinetics in Chemical Engineering."

Professor Drew is a native of Wayland, Mass., and holds a B.S. and an M.S. degree from Massachusetts Institute of Technology.

He has taught at M.I.T., Drexel Institute, and was a chemical engineer in the experimental station of the Du Pont Co., Inc. He became associate professor of chemical engineering at Columbia in 1940; was made a full professor in 1945; and head of the department in 1948. During World War II, Professor Drew was director of engineering at the S.A.M. Laboratories, the Columbia section of the Manhattan Project (Atomic Energy Commission).

In the latter part of the war he was responsible for designing the heat-removal systems for the nuclear reactors.

He is now a consultant for the Brookhaven National Laboratory and is also chairman of its Engineering Advisory Committee. Other activities in the chemical engineering field at present include that of consultant for the Du Pont Co. in the design of the Savannah River A.E.C. development.

Professor Drew won the William H. Walker Award of the A.I.Ch.E. in 1937, for his work on, "The Dropwise Condensation of Steam."

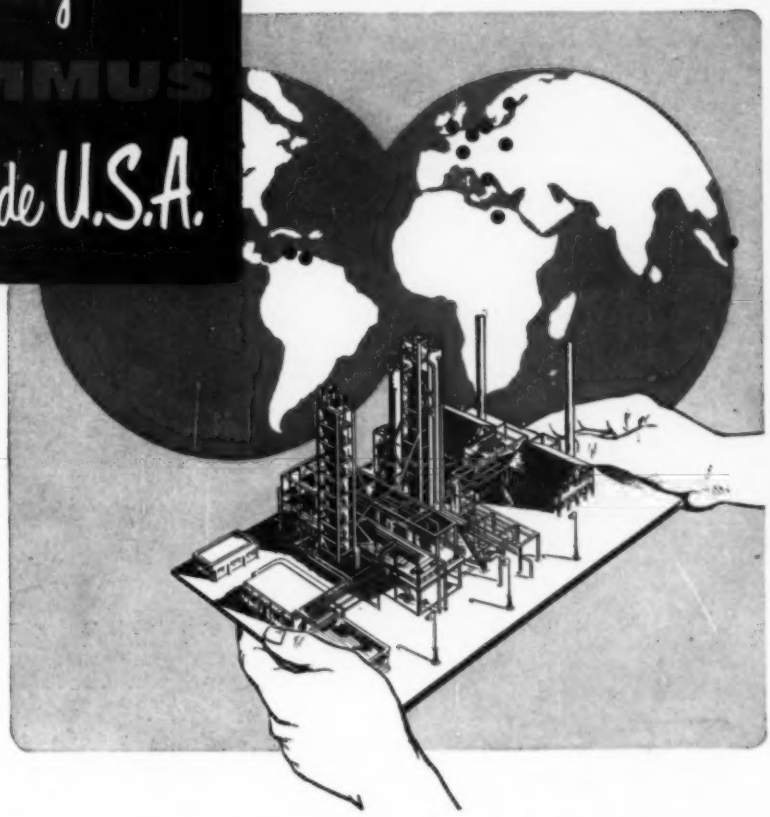
He is a member of A.S.M.E., A.C.S., A.S.E.E., Army Ordnance Assoc., and a fellow of the New York Academy of Science.

TRANSACTIONS—THE INSTITUTION OF CHEMICAL ENGINEERS

Word has just been received from The Institution of Chemical Engineers (London) of a change in publication policy, starting with its Transactions for the year 1951. The plan as it now stands is to publish the 1951 Transactions in three parts: Part One to be available shortly, Part Two in January, 1952, and Part Three in April, 1952. The 1952 Transactions will be published in four parts and their publication completed in April, 1953.

This change in publication policy will make the Transactions of The Institution available at a much earlier date than they have been in the past. Institute members will be offered a reduced subscription rate and further information regarding this will be sent to all members about Nov. 15 with the 1952 dues' bill.

...by LUMMUS ...outside U.S.A.



Dots on the globe indicate countries where scores of post-war refinery and chemical projects carry the seal "... by Lummus".

All classes of Lummus service—process design, engineering, construction—are represented in these overseas projects. In a number of cases, combined projects constitute complete oil refineries.

Processes employed include distillation; propane deasphalting; catalytic and thermal cracking; reforming; Polyforming; coking; lube oil manufacture by solvent refining, solvent dewaxing and clay treating; and petroleum chemicals production.

Each job—whether just starting, partially completed, or on stream—bears the Lummus stamp of over-all competence and international experience.

We invite you to consult Lummus on *your* next project.

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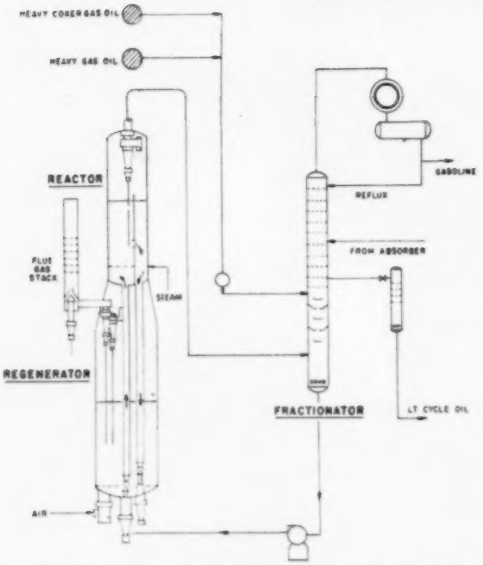
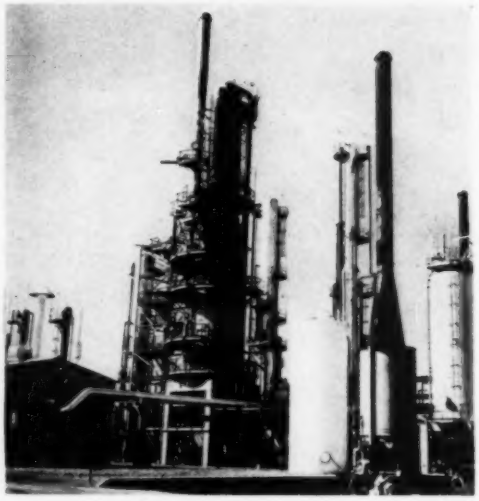
DESIGNING ENGINEERS AND CONSTRUCTORS FOR THE PETROLEUM AND CHEMICAL INDUSTRIES

ORTHOFLOW CRACKER FOR CANADA





Above is a picture and diagram of the construction and internal details of the Orthoflow Fluid catalytic cracker, recently put on stream at Edmonton, Canada, for the British American Oil Co., Ltd., by The M. W. Kellogg Co. A small unit, the Orthoflow is designed to process approximately 2,000 bbl./day.

The Orthoflow is a single-vessel unit containing chambers for reaction, regeneration and stripping with straight-line catalyst carrier lines within the vessel. The lines will be less subject to erosion by high velocity catalyst than bent lines. Another economy is the elimination of expansion joints in the lines.

To absorb thermal expansion of the carrier lines, Kellogg spring-loaded the catalyst-and-oil-flow control valves. The valves themselves are a deviation from the slide valves which control the flow in the majority of Fluid crackers. In essence, they are large needle valves through which catalyst and oil are introduced to the carrier line leading to the reactor. Hot catalyst from the bottom of the regenerator is drawn into the line by the pressure drop created by high velocity oil passing through the valve.

NEWS

(*Continued from page 24*)

HEYDEN, SHAWINIGAN FORM NEW COMPANY

Heyden Chemical Corp. and Shawinigan Chemicals Ltd., subsidiary of the Shawinigan Water and Power Company of Montreal, have decided to form a new Canadian company which will manufacture initially formaldehyde and pentaerythritol, according to a joint announcement by John P. Remensnyder, president of Heyden Chemical Corp., and V. G. Bartram, president of Shawinigan Chemicals, Ltd.

Heyden and Shawinigan will share evenly in stock ownership of the new company which will construct a two-million-dollar plant at a site to be selected in Canada, with annual capacity for production of 30 million pounds of formaldehyde and three million pounds of pentaerythritol, a solid alcohol made from formaldehyde and the base for weather-resistant paints, varnishes, and resins. Production of additional chemical products is planned later.

CHEMICAL CORPS SEEKS PERSONNEL

The Army Chemical Corps has announced that it is seeking mechanical, chemical, and bacteriological engineers for employment in the Baltimore area at salaries ranging from $4600 to $8800 per year. Application for these positions should be filed with the Board of U. S. Civil Service Examiners, Army Chemical Center, Md.

The Headquarters of the Army Chemical Corps Engineering Agency is located at Army Chemical Center, Md., 20 miles north of Baltimore. The agency directs all Chemical Corps engineering activities.

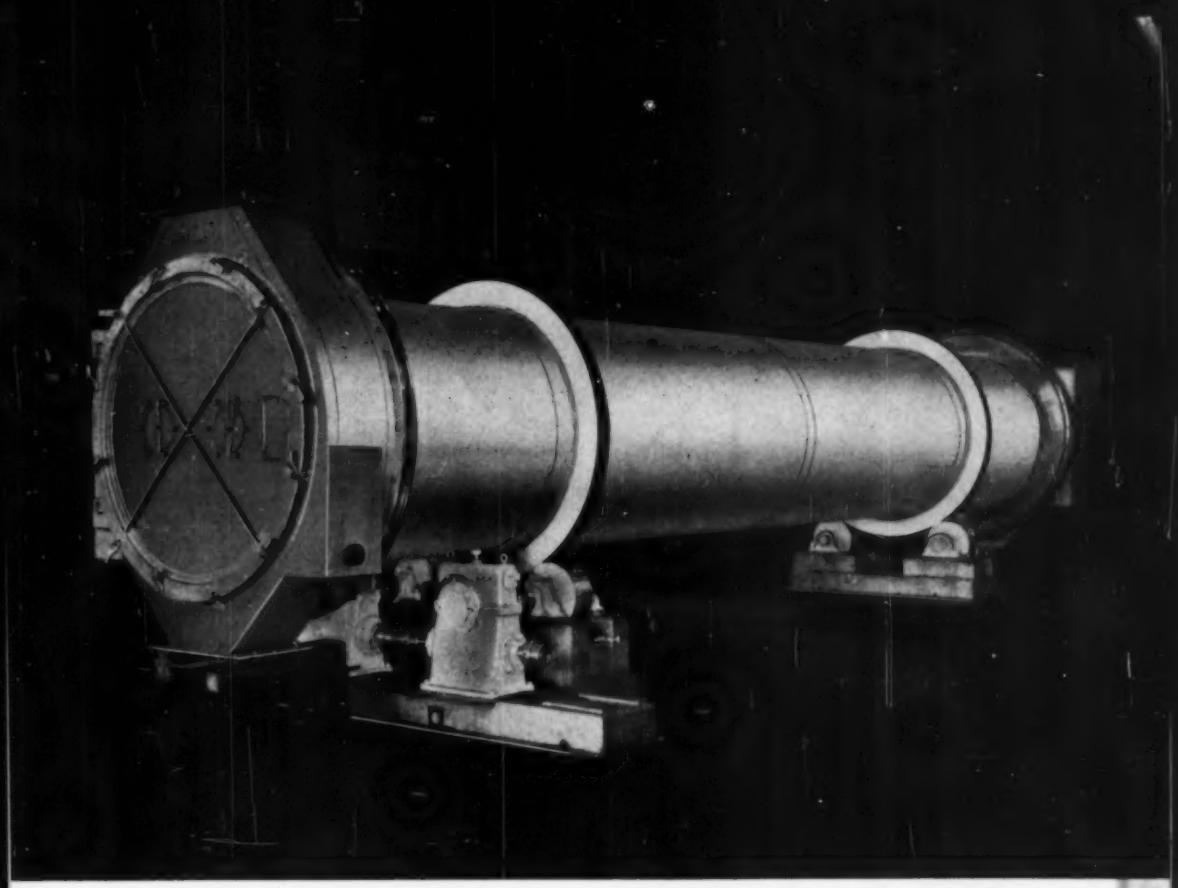
Appointments for interviews will be made on the basis of information furnished on Federal Personnel Form No. 57. Persons interested in the positions should contact the Civil Service Board at the Chemical Center. There will be no formal written examinations for the positions, and there is no closing date for filing.

E.J.C. JOINS U.P.A.D.I.

Engineers Joint Council recently received the approval of a majority of its member bodies, namely, A.I.Ch.E., A.S.C.E., A.I.M.E., A.S.M.E., and A.I.E.E., to become an adherent of Union Panamericana de Asociaciones de Ingenieros, known as UPADI. E.J.C. participatad in the first steps of this organization's formation taken at Rio de Janeiro in July, 1949, when the representatives of engineering societies in the Americas met in a First Pan American Engineering Congress. UPADI held its first meeting in Havana in 1951 and agreed on a draft constitution which was to govern until the next meeting to be held in New Orleans now scheduled for the late summer of 1952.

The United States is one of nine nations designated to furnish a member to the first UPADI board of directors. James M. Todd has been appointed U. S. member of the UPADI board and S. L. Tyler, Secretary of American Institute of Chemical Engineers, will represent the U. S. on the permanent UPADI committee on constitution and bylaws.

(*More News on page 28*)



It keeps the RED from the WHITE and the BLUE

FABRICATED entirely of stainless steel, the 60" diameter by 32'0" dryer pictured above is probably one of the most costly direct heat rotary dryers (of its size and weight) ever built.

The entire interior, including the flights, is polished to a mirror like satin finish, and is easily accessible for thorough cleaning. Both the feed and discharge breechings, complete with their air-tight inspection doors and labyrinth seals, are also stainless steel, polished to a satin finish on the inside.

To assure a satisfactory product free from contamination, the air used in drying is first filtered through an air filter. This air on leaving the dryer goes to a cyclone dust collector of Bartlett-Snow's special high efficiency design. This equipment also is of stainless steel construction, and polished, and all parts of the equipment are enameled on the outside as a further aid in preventing contamination.

Engineered, designed, and completely fabricated in our extensive machine and structural shops, this dryer—built to prevent any possible mixing and spoilage, between different batches—and colors—of a well known plastic, reflects the ability of Bartlett-Snow engineers to meet even the most exacting heat engineering problems.

Our wide experience and highly technical, mathematical approach to every dryer problem is your assurance that the diameter and length of the shell, the pitch, the rate of feed, the time of passage, the method of firing, and all other specifications of the equipment recommended will be exactly suited to your particular and individual requirements. Send for Bulletin No. 89,—and let the Bartlett-Snow heat engineers work with you on your next heat engineering problem. The C. O. Bartlett & Snow Company, Cleveland 5, Ohio. Engineering representatives in New York, Baltimore, Detroit and Chicago.

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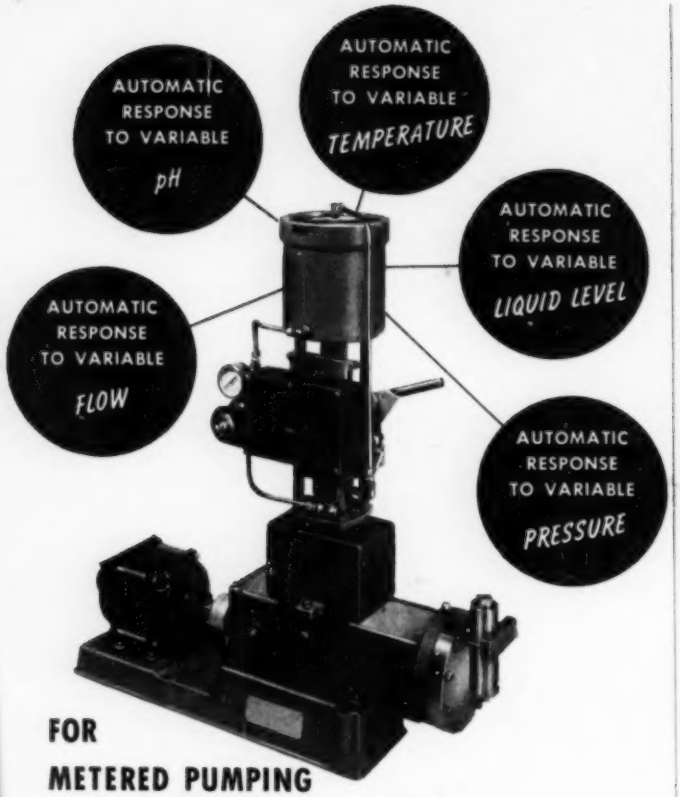

AUTOMATIC RESPONSE TO VARIABLE pH
AUTOMATIC RESPONSE TO VARIABLE TEMPERATURE
AUTOMATIC RESPONSE TO VARIABLE LIQUID LEVEL
AUTOMATIC RESPONSE TO VARIABLE FLOW
AUTOMATIC RESPONSE TO VARIABLE PRESSURE

FOR METERED PUMPING

in Automatic Response

TO VARIABLE PROCESS DEMAND

The Lapp Pulsafeeder is applicable to all pneumatic or electro-pneumatic instrumentation. Operates at constant pumping speed—variable flow results from variation only in piston stroke length, controlled by the pneumatic cylinder. It is a positive displacement pump, which operates *without stuffing box or running seal.* Its hydraulically-balanced diaphragm acts as a floating partition which isolates chemical being pumped from pump parts—to protect against contamination of product or equipment. Entire mechanism is inherently explosion-proof.

WRITE for complete description and specifications. Lapp Insulator Co., Inc., Process Equipment Division, 453 Maple St., LeRoy, N. Y.

Lapp *Auto-Pneumatic*

PULSAFEEDER

PISTON-DIAPHRAGM CHEMICAL PROPORTIONING PUMP

NEWS

(Continued from page 26)

E.C.P.D. TO MEET IN BOSTON

Engineers' Council for Professional Development, a conference of engineering bodies organized to enhance the professional status of engineers, will hold its annual meeting Oct. 19 and 20 at the Hotel Statler, Boston, Mass. Harry S. Rogers, president of Polytechnic Institute of Brooklyn, will preside over the two-day meeting which begins with registration at 9 o'clock Friday morning.

Three sessions are planned, beginning on Friday afternoon with a report of the E.C.P.D. Training Committee under the direction of A. C. Monteith, chairman, followed by a panel discussion on the implementation of the six-point professional training program presented to E.C.P.D. at the annual meeting in Cleveland last year. Actual methods of introduction of each of the main points of the program to typically urban areas will be described by engineers.

The second session will be held Saturday morning and will feature a report of the Guidance Committee by Willis F. Thompson, chairman, followed by another panel discussion on the high school viewpoint of guidance. Speaking then will be Frederick A. Small, principal, School Department, Norwell, Mass.; Thomas D. Ginn, director of vocational guidance, Boston School Committee; and Aaron Fink, guidance director, School Department, Stoughton, Mass.

The third and concluding session will be held Saturday afternoon and will hear reports from other E.C.P.D. Committees.

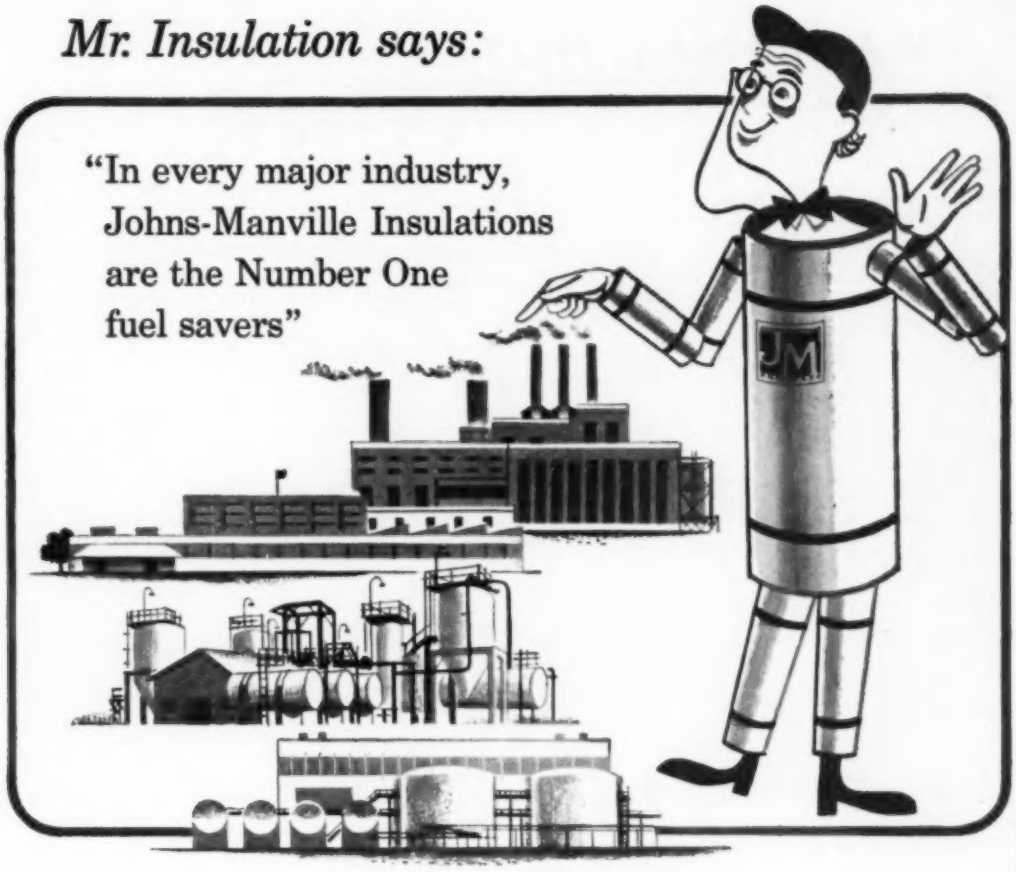
Luncheon and dinner speakers are planned and on Friday at luncheon, Harold B. Richmond, chairman of the board, General Radio Co., Cambridge, Mass., will talk on "Observations on Cooperative Course Training as Viewed by an Employer." Saturday's luncheon will be addressed by James Rhyne Killian, Jr., president, M.I.T., on "Student Counseling in an Engineering School."

The annual dinner on Friday night will feature two talks, one by H. S. Rogers on "Service, Motivation and Support"; the other by Eugen W. O'Brien, vice-president of W. R. C. Smith Publishing Co., Atlanta, "What Junior Engineers Find Important."

Visitors will be welcomed to all events. On Sunday friends of the late Dugald C. Jackson, former head of the department of electrical engineering at M.I.T. and past president of the American Institute of Electrical Engineers, and a long-time member and leader of E.C.P.D., will have an opportunity to honor his memory at a Memorial Service Oct. 21 at 4 P.M. at the First Congregational Church in Cambridge, 11 Garden Street.

(More News on page 32)

Mr. Insulation says:



IN PLANTS with stacks or stills, tanks or towers, the story is the same—"Insulation by Johns-Manville" means the maximum return on your investment. There are two good reasons for this:

1. YOU GET THE RIGHT *MATERIALS*—From asbestos and other selected raw materials, Johns-Manville manufactures a wide range of industrial insulations for service from 400F below zero to 3000F above. If you need engineering advice in determining exactly the right one for your job, you can get experienced and authoritative assistance from your local Johns-Manville insulation engineer.

2. YOU GET THE RIGHT *APPLICATION*—Even correctly selected insulation needs proper application to permit it to serve at peak efficiency through the years. Here you can always count on J-M Insulation Contractors and their highly skilled mechanics. These organizations are trained in Johns-Manville application methods, and have generations of insulation experience behind them.

Why not get the complete picture? Call on insulation headquarters when your next job is in the planning stage. Just write Johns-Manville, Box 290, New York 16, N. Y.

Johns-Manville *first in* **INSULATIONS**

MARGINAL NOTES

News of Books of Interest to Chemical Engineers

All About Beets

Beet Sugar Technology. R. A. McGinnis. Reinhold Publishing Corporation, New York, N. Y. (1951) 574 pp. $10.00.

Reviewed by Erman A. Pearson, Assistant Professor of Sanitary Engineering, University of California, Berkeley, Calif.

THIS book is a creditable presentation of the varied aspects of beet sugar technology and constitutes an important addition to the chemical engineering literature in a field of endeavor vital to the economy of the United States. Mr. McGinnis states that the book is addressed to three groups of readers: namely, to the beet sugar factory technicians for instructional and reference purposes; to the beet sugar growers and others connected with or interested in the industry for general information; and to professional engineers, chemists, and technological college students as a reference manual.

The author, as editor of a manuscript representing the contributions of 29 authors, has done a remarkable job in organizing and establishing continuity in the text. Some chapters are slightly repetitious; however, this is to be expected in a book compiled by so large a number of contributors. The treatment of the subject matter in the text oscillates from a scientific to an engineering or technical point of view interspersed with considerable general information for the nontechnical reader; however, this too is somewhat necessitated by the broad aim and scope of the book.

The material is assembled under the following general headings: history of the industry, growing sugar beets with considerable emphasis on the agricultural aspects; processing, starting with the preliminaries to processing and including diffusion, fundamental chemistry of juice purification, first and second carbonation, adsorbing and absorbing agents, evaporation, and crystallization; and a general discussion of packaging and storage, special sugar products, lime and carbon dioxide production, the saccharate processes, pulp drying, waste disposal and by-product recovery; and appendices including a glossary of technical terms, samples of processor-grower contracts and a list of books written since 1911 relating to sugar beet agriculture and beet sugar manufacture.

The subject matter is indexed and the text is replete with references for the worker seeking additional information on the numerous subjects that are only briefly reviewed.

This is the first comprehensive book in English on the manufacture of beet sugar in 44 years and is recommended as a general reference work to anyone associated with the industry in an administrative, operational or technical capacity.

An Addition to Chemist's Library

The Vitamin B Complex. F. A. Robinson. John Wiley & Sons, Inc., New York. (1951) xi + 688 pp. $9.00.

Reviewed by Leo A. Flexser, Group Chief, Manufacturing and Development Department, Hoffmann-La Roche, Inc., Nutley, N. J.

WITH the appearance of "The Vitamin B Complex" by Mr. Robinson, deputy research director of the well-known British firm of Allen and Hanburys, Ltd., we have a text which fills many of the gaps that have opened in the literature of this rapidly expanding field during the past few years. As such, it is a valuable contribution even though already slightly out of date on the recent developments in folic acid and vitamin B_{12}.

The book contains an exhaustive survey of each of the B vitamins: namely, thiamine, riboflavin, niacin, pyridoxine, pantothenic acid, biotin, folic acid, and B_{12}. Also discussed are such substances as *p*-amino-benzoic acid, inositol, choline, and other factors which are not always considered to be vitamins but whose inclusion is entirely appropriate and desirable. Each of these substances has devoted to it some 15 to 20 major sections bearing such titles as Isolation, Chemical Constitution, Synthesis, Properties, Estimation, Occurrence, Deficiency in Animals, Deficiency in Man, Metabolism, Requirements, Pharmacological Action, Analogues, etc.

Insofar as completeness of treatment is concerned, the book in general merits approval. Unfortunately, the author has written mainly for the English rather than the American reader. Thus, references are given primarily to British rather than to United States patents, even in such cases as thiamine which was, in fact, first synthesized and patented in the United States. One could also point out inadequacies and errors concerning the industrial synthesis of several of the vitamins, but these are, perhaps, pardonable in view of the paucity of reliable information.

The book will be a valuable addition to the library of the chemist or biochemist interested in vitamins. Its utility to the chemical engineer will necessarily be limited.

Petroleum Facts and Figures. Ninth Edition, 1950. American Petroleum Institute, New York, N. Y. (1951) 491 pp. $2.50.

CHARTS, graphs and other illustrations were held to a minimum in this new edition, the first since 1947, to provide maximum leeway for new tables and additional statistics. This edition stresses long-term series of comparable tables wherever possible on utilization, production, refining, marketing, transportation, prices and taxation. Statistics for the many tables were collected and compiled by the research section of the Department of Information, with the assistance of other divisions of the Institute, government agencies, oil company authorities and independent oil experts. Under the title of "General" the book includes statistics on labor; safety; fire loss; finance; world crude oil production by regions and countries, and miscellaneous statistics. The statistics are very interesting. For instance one learns that in 1949 we produced 961-odd million barrels of motor fuel, which is more than ten times the production in 1918.

In the miscellaneous department one comes across other startling facts, such as the U. S. has 70% of the world's motor vehicles, that the average octane rating of motor fuel for last year was 78 for the regular and 82 for the premium, plus various weight conversions, running conversions, and other interesting data on petroleum products. This issue also contains historical information on the A.P.I., its organization, and purposes and lists officers and directors for 1951.

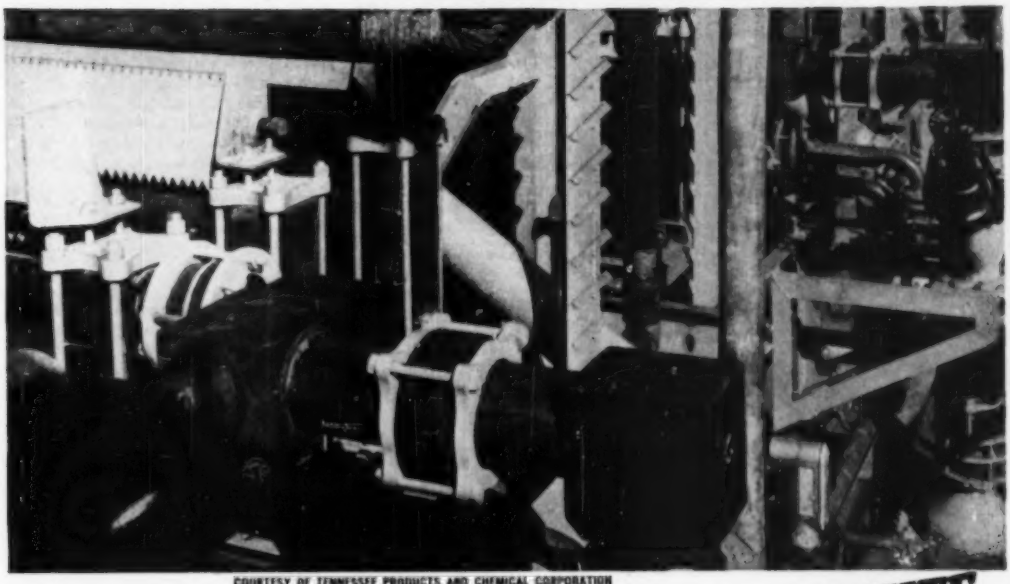
Books Received

Fundamental Principles of Physical Chemistry. Revised edition. C. F. Prutton and S. H. Maron. The MacMillan Co., New York, N. Y. (1951) 803 pp. $5.75.

Radiochemical Studies: The Fission Products. 3 vols. Book 1: Parts I-IV—Papers 1 to 52, 516 pp. Book 2: Part V—Papers 53 to 199, 1315 pp. Book 3: Parts V (continued), VI-VIII—Papers 200 to 336, 2086 pp. Edited by C. D. Coryell and Nathan Sugarman. McGraw-Hill Book Co., Inc., New York. (1951) $18.50.

QUESTION:

What pipe handles ALL corrosives?



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ANSWER:

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For conveying, processing and storing *all* types of corrosive liquids and gases, hot or cold, whole systems are now being built 100% with "Karbate" impervious graphite pipe, fittings, valves, pumps, heat exchangers, towers and other units needed. Here are the reasons:

★ **When you use "Karbate" impervious graphite, you don't need to use a variety of corrosion-resistant materials.**

★ **"Karbate" impervious graphite pipe keeps production rolling. You don't have costly "down time" because of corrosion in the system. There is no loss of product.**

PLUS FACTORS OF "KARBATE" IMPERVIOUS GRAPHITE PIPE:

- Both acid resistant and alkali resistant
- Light in weight, easy to machine and assemble
- No metallic contamination of product
- Immune to thermal shock

OTHER NATIONAL CARBON PRODUCTS

The terms "Karbate" and "Eveready" are registered trade-marks of Union Carbide and Carbon Corporation

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WHY BUY THIS ITEM?

It gives you *double the usable light!*
It's *leakproof!*
It can't *stick, swell or jam!*
It delivers the *whitest, brightest light!*

It's the "Eveready" No. 1050 flashlight battery, made with the zinc electrode inside a carbon jacket – just the reverse of every other battery on the market.

Test it and . . . you always buy it!

HEAT EXCHANGERS • PUMPS • VALVES • PIPING • TOWERS • TOWER PACKING • BUBBLE CAPS • BRICK • STRUCTURAL CARBON • SULPHURIC ACID CUTTERS • HYDROCHLORIC ACID ABSORBERS

PRODUCTS OF METHYLATION

IT'S *Easy* AND *Economical* TO METHYLATE WITH ANSUL METHYL CHLORIDE

You can save money by using ANSUL METHYL CHLORIDE in the methylation of amines, phenols, alcohols, hydrocarbons, etc. Based on CH_3-groups available for reaction, other methylating agents are considerably more expensive than Ansul Methyl Chloride.

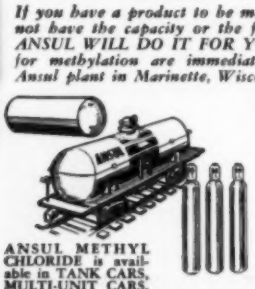
Also Ansul Methyl Chloride can be conveniently weighed or metered for positive control of the methylation process. When methyl chloride is used for methylation, no pumps are necessary for transferring it. Methyl chloride, a liquefied gas, provides its own pressure and facilitates handling greatly.

If you have a product to be methylated and you do not have the capacity or the facilities to handle it, ANSUL WILL DO IT FOR YOU. Ample facilities for methylation are immediately available at the Ansul plant in Marinette, Wisconsin.

Write for your copy of Bulletin No. 1003 an Ansul Technical Staff publication.

If you wish specific information regarding methylation processes, products of methylation or products you wish methylated give particulars in your letter. Our Technical Staff will willingly cooperate with you.

For your SULFUR DIOXIDE requirements use ANSUL SULFUR DIOXIDE.



ANSUL METHYL CHLORIDE is available in TANK CARS, MULTI-UNIT CARS, DRUMS and CYLINDERS.

LIQUID Methyl Chloride by ANSUL

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ALSO MANUFACTURERS OF REFRIGERANTS • REFRIGERATION PRODUCTS • DRY CHEMICAL FIRE EXTINGUISHERS

NEWS

(*Continued from page 28*)

SEES THREE MILLION MORE TONS OF SULFUR

J. C. Carrington, assistant to the president of Freeport Sulphur Co., speaker before the American Institute of Mining and Metallurgical Engineers, listed 46 projects, including 24 in the United States, which will produce sulfur in different forms and which are expected to yield, on the basis of present estimates, approximately 3,000,000 long tons of the mineral by the end of 1953. Government sources have placed the world shortage of sulfur at 1,000,000 tons annually.

He listed the projects as follows:

1. Twenty-four projects in the U. S. Four are new brimstone mines and one the expansion of an existing mine with a total anticipated new production of 1,200,000 long tons of sulfur. One of the new mines already is in operation. A second is scheduled to start production in the first half of 1952, another by the end of the same year, and the fourth, the newly discovered deposit at Garden Island Bay in Louisiana which Freeport will develop, by the end of 1953. Other projects are expected to add 350,000 tons of sulfur from natural gas, petroleum refinery gas, smelter gases, pyrites and low-grade surface deposits of sulfur.
2. Seven projects in Canada and four in Mexico with an estimated new production of 500,000 tons of sulfur in all forms. A plant to mine sulfur by the Frasch process will be built at a salt dome deposit on the Mexican Isthmus of Tehuantepec.
3. Expansion of pyrites production in Spain by increasing the annual rate of output at mines in Huelva Province by an amount estimated to be the equivalent of 400,000 long tons of sulfur.
4. Increases planned by Italy in the output of native sulfur amounting to about 250,000 tons yearly.
5. Projects in Great Britain, Australia, Colombia, Japan, Norway and Ecuador to obtain more sulfur in various forms from anhydrite refinery gases, pyrites and native deposits.

SAVE THOSE BACK ISSUES

Every so often an unprecedented demand for a particular issue, or an unexpected influx of new subscribers and members puts the editor in the embarrassing position of running out of copies of Chemical Engineering Progress. This has happened several times in our short history and if members have copies of any of the following issues, we would be glad to purchase them. The issues which we need and for which we will pay 75 cents each, are: January, 1949; October and November, 1950, and February and March issues of 1951.

All these issues were overprinted to a great extent, but because of features and other demands, single copy sales, etc., they were completely exhausted in a short time.

A.E.C. EXTENDS G. E. CONTRACT

The Atomic Energy Commission has extended for five years the General Electric Co.'s contract for operation of the Hanford (Wash.) Works and the Knolls Atomic Power Laboratory, Schenectady, N. Y.

The extended contract will run until June 30, 1956. General Electric has operated the Hanford plutonium plant for the government since September, 1946. All costs incurred by General Electric in carrying out its contract are reimbursed from government funds. The fee included in the extension is the one dollar stipulated in the original contract. The contract extension also provides for re-negotiation of the one-dollar fee at some time in the future if this seems desirable.

MONSANTO TO RUN CHEM. CORP PLANT

Monsanto Chemical Co. announced last month it has signed a letter contract with the Chemical Corps of the Department of the Army for the operation and maintenance of the Chemical Corps' chlorine caustic plant at Muscle Shoals, Ala. The plant will produce chlorine by the De Nora mercury-cell process which Monsanto has adapted for use in the United States and Canada. The company announced it will initiate negotiations immediately with the Chemical Corps for a formal contract.

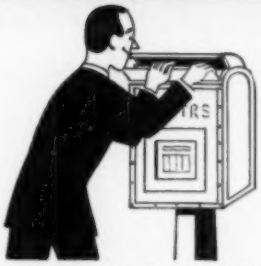
J. H. Zwemer of Sheffield, Ala., who has been project manager during construction, will be plant manager for Monsanto.

NEW ASSOCIATION OF SALES ENGINEERS

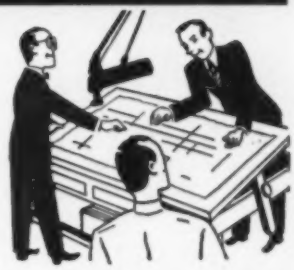
A group of sales engineers in the chemical equipment field, who have been meeting informally in New York since December, 1950, have organized the Chemical Equipment Sales Engineers Association. First officers elected were K. S. Valentine, Patterson Foundry & Machine Co., president; J. Robert Powers, Baker-Perkins Co., vice-president, and Laurence C. Johnston, Brown Fintube Co., secretary-treasurer. Mr. Valentine and Johnston are members of A.I.Ch.E. Membership at present is limited to qualified sales engineers in the New York district. Luncheon meetings are held each month at the Chemists' Club for informal discussion of current problems or for talks by guest speakers. At a recent meeting C. H. Chilton, a member of A.I.Ch.E. and associated with Chemical Engineering, talked on cost estimation.

(More News on page 48)

HOW TO SAVE TIME and MONEY IN BUYING DRYING EQUIPMENT

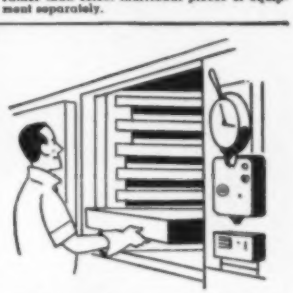


1 The first step in saving time and money in buying drying equipment is to become thoroughly acquainted with the Proctor approach to solving drying problems . . . by writing for Bulletins 343 and 361. These bulletins explain the way Proctor engineers tackle your problem.

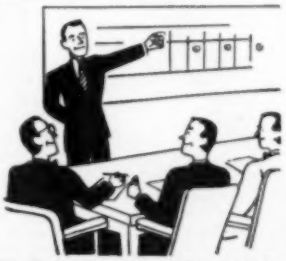


2 With this information as a background, consider your equipment needs just as soon as you know that drying is to be a part of your process. View that process as a whole, rather than select individual pieces of equipment separately.

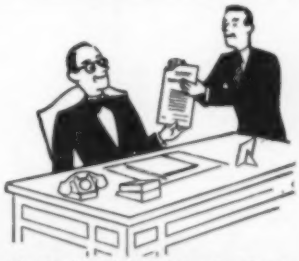
3 Pick up your telephone—or if that is inconvenient—dictate a letter outlining your problem. A Proctor engineer will be in touch with you to discuss it in detail and arrange for test work.



4 Tests will be conducted either in the Proctor laboratory—or if the nature of your product demands, in your own plant. The results will serve as a guide in making preliminary plans for equipment.



5 On the basis of the test work Proctor engineers will present recommendations on the size and type of equipment required to meet your needs. This equipment will be planned to serve as a major link in the over-all process.



6 Based on all of the test work and engineering that has been done—your sales contract will guarantee the performance of your Proctor system. You save time, money and are assured equipment that is "job-engineered" for you.

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DATA SERVICE

BULLETINS

1 • WATER CONDITIONING. For conditioning water for boiler feed, process work and other industrial needs Penfield Mfg. Co., Inc., explanatory bulletin on its water-conditioning equipment. Features advantages of planning processes for water purity required. Illustrations of monobed demineralizers, four-column demineralizers, silica removers, water softeners, degasifiers, deaerators, filters, and illustrations of packaged units.

2 • MILLS AND MIXERS. Description of Troy Engine & Machine Co. line of roller mills, colloid mills, and angular mixers. Roller mill features rigid roll support with less deflection and consequent greater grinding surface. Operator can set apron and feed rolls simultaneously, keeping them parallel. Graduated dial makes possible duplication of previous run. Angular mixer features an agitator and turntable with separate drives permitting independent and combined operation. Illustrated with sizes, types, weights, machine and motor sizes, etc.

3 • FIRE EXTINGUISHER GUIDE. To help employees select correct fire extinguisher for wood, inflammable liquid, or electric fires, Randolph Labs, Inc., have issued a chart for posting at strategic plant locations.

4 • MATERIALS-HANDLING. Engineering booklet on how to cut costs in materials-handling by the Yale & Towne Mfg. Co. Revision previous issue prepared under Dr. V. S. Karabasz of the Wharton School, University of Pennsylvania. Well illustrated, shows the various kinds of equipment, the fundamentals in handling materials with automatic equipment, shows how to use vis-o-graf—a step-by-step analysis of handling problems, and illustrates the various units available for materials handling.

5 • PROCESS INDUSTRIES EQUIPMENT. A catalog of the special fabrication services of Black, Sivalls & Bryson, Inc., to the process industries. Describes equipment they can make, heaters, storage tanks, control valves, liquid level controllers, drainers, etc., for the petroleum and chemical process industries. Illustrated, specifications given, etc.

6 • STORAGE HEATERS. For plants with varying hot water demand rates Whitlock Mfg. Co. describes in new bulletin various storage heaters where requirements for hot water are intermittent. Well-illustrated, shows construction, materials used, dimensions of all equipment, pressure tables, methods of selecting the proper size heaters, conversion factors. Typical layouts.

7 • INSTRUMENT TUBING. Samuel Moore & Co. brochure on a corrosion-resistant instrument tubing for use in process industries. Tubing coated with plastic, either polyethylene or vinyl over a core of steel, aluminum, or copper. Low cost for use in installations where fumes or sprays are corroding control lines. Straight lengths or 60 ft. coils, ¼ to ½ in. O.D.

8 • STORAGE TANKS. Diaflote, a new development in storage tanks, in a bulletin from Hammond Iron Works. Features a synthetic film membrane inside the tank resting on the surface of the stored liquid. This acts as a barrier between the liquid and air preventing formation of vapors. Complete details of construction, use, etc. Of interest to the petroleum and process fields.

9 • CONDENSATE UNIT. For installations with minimum floor space, Roy E. Roth Co. has newly designed and described a vertical condensate return unit for handling water to 210° F. without vapor lock. Various sizes, contains pump, receiver, float switch, etc.

10 • DIAPHRAGM MOTOR VALVE. Minneapolis-Honeywell Regulator Co. specification sheet for air-operated diaphragm motor valve. Contains description of the units, sizes, construction, body material, pressure ratings, as well as dimensions. Chart shows flow-lift characteristic curves for on-off and narrow band proportional control valves.

11 • INDUSTRIAL MIXERS. Grenier & Co. bulletin on industrial mixers and laboratory stirrers. Features propeller type for production, pilot plant and laboratory, portable as well as side-entering mixers, also flange-mounted. Laboratory models explosion-proof and air driven.

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12 • *WEATHERPROOF MOTOR STARTERS.* For the installation which requires waterproof or explosion-proof starter for motors in hazardous locations, Arrow-Hart & Hegeman Electric Co. details in a bulletin its line meeting these requirements. Underwriter approved. Information on dimensions, weights, ratings, wiring diagrams, etc.

14 • *PROCESS INDUSTRY BULLETIN.* Infilco, Inc., in a new bulletin for the process industries tells how engineering problems have been solved in paper manufacture, pulp mills, oil producers, plating wastes, chemical plants, oil refineries, etc. Problems are typical and equipment used illustrative of type made by Infilco. Problems range from clean up of water from plating, corrosion prevention in cracking stills, and removal of products causing scaling on heat exchangers.

15 • *WELDING FITTINGS.* A Tri-Clover Machine Co. illustrated catalog of stainless steel welding fittings. Used with light gauge pipe. Bulletin contains about every type of fitting available, plus miscellaneous tables of technical data.

16 • *THERMOMETERS AND GAUGES.* The Electric Auto-Lite Co. with a descriptive bulletin and information on thermometers, dial-indicating thermometers, recording gauges for pressure and vacuum, transformer thermometers and freezer thermometers. Details of construction of each are shown. Illustrated with data and standard construction details.

17 • *HEAT TRANSFER.* For the chemical engineer or designer concerned with heat transfer through metallic walls International Nickel Co., Inc., has a booklet which is well illustrated and begins with the basic theory of heat transfer and brings it slowly through its mathematical form to design considerations. Typical illustrations are given as well as 12 charts on the over-all heat-transfer coefficients of various types of metals.

18 • *TACHOMETER.* For measuring speeds of two to ten processing machines with a single fixed indicator, Metron Instrument Co. has described in a technical data sheet a new Tachometer indicator. Fifty-one standard full-scale range markings available permitting direct measurements from any speed between $\frac{1}{10}$ to 100,000 rev./min.

19 • *INSULATING CEMENT.* Folder from Baldwin-Hill Co. on a new finishing cement which insulates. Temperature ranges up to 1700° F.

20 • *SYNTHETIC FILTER CLOTHS.* Samples of six different synthetic filter cloths, plus descriptive material comparing them, a selector chart listing best materials for typical corrosive chemicals are in a new folder released by Filtration Engineers, Inc. Covers two types of nylon, two types of Orlon, Dynel and Vinyon-N.

21 • *FOAMGLAS INSULATION.* Foamglas, a rigid cellular glass insulating material and its application to industrial insulation, is explained by Pittsburgh-Corning Corp. For temperatures ranging from −350° F. to 212° F. Data book gives illustrations of use and installation, sizes and thickness, application data for pipes, fittings, processing equipment, shapes available, typical application data, properties, thermal conductivity, plus characteristics and sources of accessory materials.

22 • *STAINLESS TEMPERATURE CHART.* A Carpenter Steel Co. 1-page data table, for engineers using operations at elevated temperature, on high temperature properties of stainless steel tubing. Seven different steels are listed from temperatures ranging to 1800° F. Gives tensile strength, stress, embrittling grain, thermal conductivity, specific heat and resistance to intergranular carbide precipitation.

23 • *ENTRAINMENT REMOVER.* For use inside steam drums, packed towers, bubble-cap towers, deodorizers, stills, evaporators, etc. V. D. Anderson Co. has designed and published a bulletin on new internal entrainment remover—Hi-eF purifier. Delivers vapor with 1 ppm or less of total solids. For application to chemicals, petroleum, gas, power, etc., industries. Purifier uses rotary flow of gases to stop entrained solids. Made in various materials, designed to pass through standard manholes.

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EQUIPMENT

26 • *SELF-PRIMING MOTORPUMP.* Ingersoll-Rand Co. is introducing a new line of self-priming motor pumps. Intended for suction lift. Primes itself by recirculating liquids trapped in casing. Built in sizes from ¼ to 25 hp. Capacities up to 800 gals./min. Heads up to 180 ft.

27 • *VIBRATORS.* For prevention of arching, hanging-up and sticking of bins, hoppers, chutes, etc., Cannon Vibrator Co. with new vibrators 1¼ in. to 4 in., weight from 9 to 140 lbs., features air-cushioned impact to eliminate metal to metal pounding.

28 • *BIN-LEVEL INDICATOR.* For level indication in large bins the Bin-Dicator Co. has designed a new model for top suspension anywhere in the bin. Specially adaptable where materials build up on walls making wall models impractical.

29 • *CORROSIVE FUME FAN.* Clarage Fan Co. is producing a corrosion-resistant fan from stainless steel for use in processing plants moving sulfuric and nitric acid fumes.

30 • *GAMMA TUBE.* Stainless steel tube designed by Tracerlab, Inc., for industrial process control procedures using gamma rays. Use as in oil pipeline inspection, liquid level gauge, etc. Fill gas used is helium. Available in special lengths.

31 • *WASHER-DEWATERER.* A small diameter, screw conveyor washer and dewaterer made by the Eagle Iron Works. For bulk-washing of loose material and delivery to stock pile, belt conveyor, or chute. Contamination is flowed over a three-sided weir, wash water being introduced through inlets in the bottom of the tub.

32 • *LIQUID-PHASE HEAT TRANSFER.* American Hydrotherm Corp. announced a line of liquid phase heat-transfer equipment to operate at temperatures up to 800° F. Uses three heat-transfer liquids—one nonflammable, noncorrosive, pumpable at room temperature and usable up to 600° F. Another pumpable at 20° F., high flame point, operating up to 650° F. The third is usable up to 800° F. All operate at atmospheric pressure. For oil, gas or coal-fired burners with capacities up to 20,000,000 B.t.u./hr.

33 • *REDUCER FITTING.* To permit different sizes of tubing to be joined easily Crawford Fitting Co. has developed the Swagelok reducer. To install tubing is merely inserted into the reducer and reducer nut tightened. The reducer is then put into a swagelock fitting which is also tightened. For instrumentation and process piping lines. Available in various metals ranging from ⅛ in. to ⅞ in. O.D.

34 • *THERMOCOMPRESSION EVAPORATOR.* For the special installation which requires pure distilled water for drinking, ice-making, processing where water is difficult to obtain, Mechanical Equipment Co. has developed thermocompression evaporators in various capacities and drives.

35 • *WEATHERPROOF CENTRIFUGAL FAN.* The Trane Co. is producing a centrifugal fan arrangement with weatherproof hood that can be installed outdoors without further protective enclosure. Wheel diameters as large as 60 in.

36 • *PURGE METER.* An acrylic plastic purge meter that eliminates packing glands and glass parts developed by Brooks Rotameter Co. Four sizes ranging from 0.01 to 54.0 gals./hr. of water and 0.2 to 216 cu.ft./air. Flow visible by spherical float in the metering tube. Safe to pressures of 100 lbs./sq.in. and temperatures to 200°.

37 • *SCRUBBER PURIFIER.* A two-stage dry scrubber purifier for delivering vapor with one part per million or less solids is newly in production by the Centrifix Corp. Features an anti-turbulent shield to effect primary stage separation.

38 • *CUSTOM SCALES.* For the chemical industry the Winslow Scale Co. offers a specialized scale-building service. Designs up to 200,000 lbs. for use with cone-shape or flat-bottom tanks with either direct reading or ticket-printing weigh beam.

39 • *CUPCASE THERMOMETER.* For the oil industry in determination of

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temperatures of oil Weston Electrical Instrument Corp. is offering a new cupcase thermometer narrow enough to fit into 2-in. gauge pipes in oil storage tanks. To A.S.T.M. specifications, thermometer has a cup capacity of 100 ml. plus other features for easy reading, etc.

40 • ***DISPERSATOR.*** An improved version of Premier Mill Corp. dispersator for mixing or blending two liquids of different specific gravities. Features a cylinder shaped like a cone revolving at high speed. Centrifugal force pushes products out through narrow slots and lower cone.

41 • ***OPPELLER PUMPS.*** For general medium pressure and temperature service up to 1000 gals./min. and 1200 lbs./sq.in. temperature to 400° F. and service on boiler feeds, oil refining, and handling liquids from propane to caustic, etc. DeLaval Steam Turbine Co. are in production with a new line of multistage opposed impeller pumps.

42 • ***MICROVAC PUMP.*** For high vacuum processing a Stokes microvac pump with a capacity of 500 cu.ft./min. and suitable for any type of vacuum processing work. Water-cooled, operates at 390 rev./min. with a 25 hp. motor.

43 • ***SPRAYWELDER.*** A powder metallizing unit for applying uniform overlays of hard facing alloys and bonding to base metals announced by the Wall Colmonoy Corp.

44 • ***REDUCTION MILLS.*** C. H. Wheeler Mfg. Co. announces a change in policy for its fine particle reduction mills formerly available on a rental basis now available for outright purchase. Mill for reduction of materials to submicron size by discharge of feed and a pressurized fluid at sonic or supersonic velocity. Available in two sizes ranging from 200 to 1800 lb. and from 400 to 6500 lb. feed rate.

45 • ***TEL-O-SET.*** A new instrument for providing sensitive control of flow, temperature, pressure and liquid level, etc., developed by the Minneapolis-Honeywell Regulator Co. Model is made in two types, fixed proportional band and adjustable proportional band and the unit operates wholly in response to pneumatic signals.

CHEMICALS

55 • ***IODINE.*** An extensive 74-page book on iodine, its properties and technical applications from the Chilean Iodine Educational Bureau, Inc. Every phase of the element is covered from history to its most advanced chemical reactions and properties. Complete literature survey and specific chapters on its uses. For instance: in organic chemistry, metallurgy, electricity, etc.

56 • ***SYNTHETIC WAX.*** To duplicate Japan wax Concord Chemical Co. is offering a new synthetic Aquawax which has Japan wax characteristics plus property of self-emulsification. For use in manufacture of plastic starches, finishing of textiles, leather, paper, thread, etc.

57 • ***GLYCERINE FACTS.*** Another edition of the well-known series, "Glycerine Facts," from the Glycerine Information Service. This particular issue covers use of glycerine as a cell preservative, agricultural aid, softener for nylon and other new and novel uses.

58 • ***ORGANIC CHEMICALS.*** Carbide & Carbon Chemicals Co. offers its 1952 edition of the book "Physical Properties of Synthetic Organic Chemicals." Sixteen pages, issued annually. A handy guide for uses of organic chemicals. Presents data on more than 300 items, 38 new. Covers alcohols, glycols, diols, plasticizers, esters, acetals, ketones, ethers, acids, nitrogen compounds, etc.

59 • ***ORTHO-NITROBIPHENYL.*** Monsanto Chemical Co. describes in a 10-page booklet ortho-nitrobiphenyl a low-cost primary plasticizer. Is compatible with a large range of resins synthetic and natural. Suggested uses in manufacture of dye stuffs, for use in polyvinyl acetate adhesives; polystyrene latex paints; cellulose acetate molding compounds, etc. Bulletin contains data on physical and chemical properties, its use as a plasticizer in most of the plastics, handling information, etc.

60 • ***BIOCHEMICAL BUFFER.*** Sigma Chemical Co. now making highly purified Tris (hydroxymethyl) aminomethane. Use as a biochemical buffer and primary standard. Compares favorably with potassium acid phthalate. Stable at elevated temperatures.

61 • ***FIRE EXTINGUISHER FLUID.*** Chlorobromomethane a fire extinguishing agent is now manufactured by Eston Chemicals, Inc. Tested by the armed forces. Effective in fighting magnesium and electrical fires. Reported to be twice as effective as carbon tetrachloride.

62 • ***TEXTILE PRESERVATIVE.*** For protection of textiles and other materials against microorganisms Dow Chemical Co. is manufacturing copper 3-phenylsalicylate. Approved by the Quartermaster. Material is applied to textiles from an aqueous solution or organic solvent.

63 • ***CAUTIC SODA.*** For users of caustic soda Pennsylvania Salt Mfg. Co. offers a new bulletin. Four sections—purchasing, handling and storage, equipment design, and technical data. Has instructions on safe handling of caustic in tank cars, for steaming and loading, etc.

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ROCHESTER MEETING

(*Continued from page 23*)

the colleges and offer the graduates from an engineering school a wage comparable to that of an expert steam fitter?"

"If industry wants engineers it must really go to bat for them," he said.

He also made several points on how the supply of engineers could be increased, among them the attraction into the engineering colleges of college freshmen, mathematics students, and junior college students.

"It would be possible," he continued, "to affect the enrollment of engineering students to be graduated in 1955 if industry really wanted to do it.

The second round-table discussion, under the supervision of W. E. Gift, superintendent of industrial relations with Tennessee Eastman Corp., featured first Prof. J. H. Boyd, consulting chemical engineer and visiting professor at Columbia University, on whether or not there should be two undergraduate curricula. Dr. Boyd stated that in his opinion, ". . . a broad training based on scientific fundamentals is the best curriculum developed to date for a chemical engineering course." "Further," he said, "the time limitation of the undergraduate curriculum does not permit thorough

J. H. Boyd, Columbia University, and consultant; L. A. Darling, Du Pont Co.

training of an undergraduate in a professional specialty. Specialization must be accomplished in the graduate school or by actual work in industry." He thought that improvements in chemical engineering curricula which could be effected, were: (a) bettering the communication powers of the engineering graduate through the written or spoken word; (b) stressing of honesty, and integrity by example and insistence; and (c) the addition of a course in common sense.

Dr. Boyd was followed by W. T. Nichols, director of the general engineering department of Monsanto Chemical Co., who presented the problem of how industry and colleges can cooperate to improve chemical engineering education. Nichols made a survey of more than 400 industrialists and college instructors, covering eleven classifications.

(*Continued on page 52*)

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CANDIDATES FOR MEMBERSHIP IN A. I. Ch. E.

The following is a list of candidates for the designated grades of membership in A.I.Ch.E. recommended for election by the Committee on Admissions.

These names are listed in accordance with Article III, Section 7, of the Constitution of A.I.Ch.E.

Objections to the election of any of these candidates from Active Members will receive careful consideration if received before November 15, 1951, at the Office of the Secretary, American Institute of Chemical Engineers, 120 East 41st., New York 17, N. Y.

APPLICANTS FOR ACTIVE MEMBERSHIP

V. E. Baikow, *New York, N. Y.*
C. R. Bender, *Atlanta, Ga.*
Wm. A. Bender, *Bloomfield, N. J.*
M. J. Bogese, *Hopewell, Va.*
Donald T. Bonney, *College Park, Md.*
John R. Borst, *Sherburne, N. Y.*
Calvin A. Campbell, *Midland, Mich.*
Thoren Priest Cook, *Bartlesville, Okla.*
Albert H. Cooper, *Lewisburg, Pa.*
B. Narayan Das, *Madras State, India*
W. T. Dickens, *Springfield, Mass.*
V. Wayne Fergus, *Sunray, Tex.*
R. E. Fiedler, *Decatur, Ill.*
Arthur W. Flynn, *Bronx, N.Y.*
William P. Given, *Levittown, N. Y.*
Luke Joseph Governale, *Baton Rouge, La.*
G. F. Gruenke, *Port Arthur, Tex.*
John R. Guala, *Pittsburgh, Pa.*
Laverne T. Heaps, *Schenectady, N. Y.*
Edward J. Helm, *Pittsburgh, Pa.*
Robert L. Irvine, *Pittsburgh, Pa.*
John R. Jacobucci, *Green River, Wyo.*
J. Clark Johnson, *Memphis, Tenn.*
Takuro Kamiya, *Kumamoto-ken, Japan*
Edward L. Kinsella, *Pittsfield, Mass.*
Wm. J. Kruppa, *Somerville, N. J.*
Hans J. Lang, *Monrovia, Calif.*
M. G. Larian, *East Lansing, Mich.*
Maurice R. Lyons, *New York, N. Y.*
Tomozo Maeda, *Kumamoto-ken, Japan*
Willard Marcy, *Phila., Pa.*
L. J. McCarthy, *Englewood, N. J.*
F. Charles Moesel, *Ann Arbor, Mich.*
M. G. Morningstar, *Cuyahoga Falls, Ohio*
Edward A. Murray, *Anderson, S. C.*
Albert E. New, *Texas City, Tex.*
Henry F. Nolting, *Munster, Ind.*
Arne O. Olson, *Pittsburgh, Pa.*
John J. O'Neill, Jr., *East Alton, Ill.*
Earl H. Osborne, *Hamilton Square, N. J.*
Laurel E. Putman, *Alexandria, Va.*
Howard F. Rase, *Madison, Wisc.*
T. A. Ruble, *Amarillo, Tex.*
Howard H. Reynolds, *Belmont, Mass.*
Grant E. Russell, *Webster Groves, Mo.*
Frederick G. Sawyer, *Stanford, Calif.*
Douglas A. Schoerke, *Philadelphia, Pa.*
George F. Sharrard, *New York, N. Y.*
Joseph H. Simons, *Gainesville, Fla.*
Robert E. Smith, *Rochester, N. Y.*
W. M. Stratford, *New York, N. Y.*
Francis M. Sullivan, *Newark, Del.*
Morris Tarnpoll, *Newark, N. J.*
Carleton L. Taylor, Jr., *East Gary, Ind.*
Richard K. Toner, *Princeton, N. J.*
Fred R. Walter, *Whiting, Ind.*
William A. Wellborn, *New Orleans, La.*
Henry A. White, Jr., *Richland, Wash.*
V. C. Williams, *Easton, Pa.*
Edgar E. Wilson, *Chadds Ford, Pa.*
Robert E. Wood, *Baton Rouge, La.*
Stanley B. Zdonik, *Brookline, Mass.*
Joseph P. Boyer, *So. Charleston, W. Va.*
William B. Cook, *Pasadena, Tex.*
Peter J. Davis, *Richland, Wash.*
Robert William Erickson, *Chicago, Ill.*
Irwin Frankel, *Argo, Ill.*
Manuel E. Fuentevilla, *Woodbury, N. J.*
John A. Hartsock, *Palo Alto, Calif.*
James H. Hoorman, *Wyandotte, Mich.*
George B. Hutchings, *China Lake, Calif.*
Gerhard M. Kahn, *Jamaica, N. Y.*
John Killian, *Sunset Beach, Calif.*
Jack B. Kirven, *Alhambra, Calif.*
Merlin M. Koenecke, *Baton Rouge, La.*
Stanley S. Kurpit, *Wilmington, Del.*
William B. Leder, *New York, N. Y.*
Elias Leventhal, *Los Angeles, Calif.*
Roland G. Lindsey, *Niagara Falls, N. Y.*
Fream B. Minton, *Richland, Wash.*
Richard H. Narwold, *Seaford, Del.*
Robert G. Norton, *Pampa, Tex.*
K. Keith Okamoto, *Austin, Tex.*
Victor M. Otakie, *Los Angeles, Calif.*
George F. Pappas, *Metuchen, N. J.*
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James David Wideman, *Jersey City, N. J.*
Henry Yakubowitz, *Los Angeles, Calif.*

APPLICANTS FOR ASSOCIATE MEMBERSHIP

William J. Dyck, *New York, N. Y.*
Laurie Jos. Molaison, *Gretna, La.*
Fred C. O'Rourke, *Baton Rouge, La.*

APPLICANTS FOR JUNIOR MEMBERSHIP

Sol Berg, *Elmhurst, N. Y.*
Lauren M. Beyard, *Hagerstown, Md.*
Frederick K. Bieri, *Red Wing, Minn.*
Wayne W. Binger, *New Kensington, Pa.*
William J. Bonine, Jr., *Havertown, Pa.*
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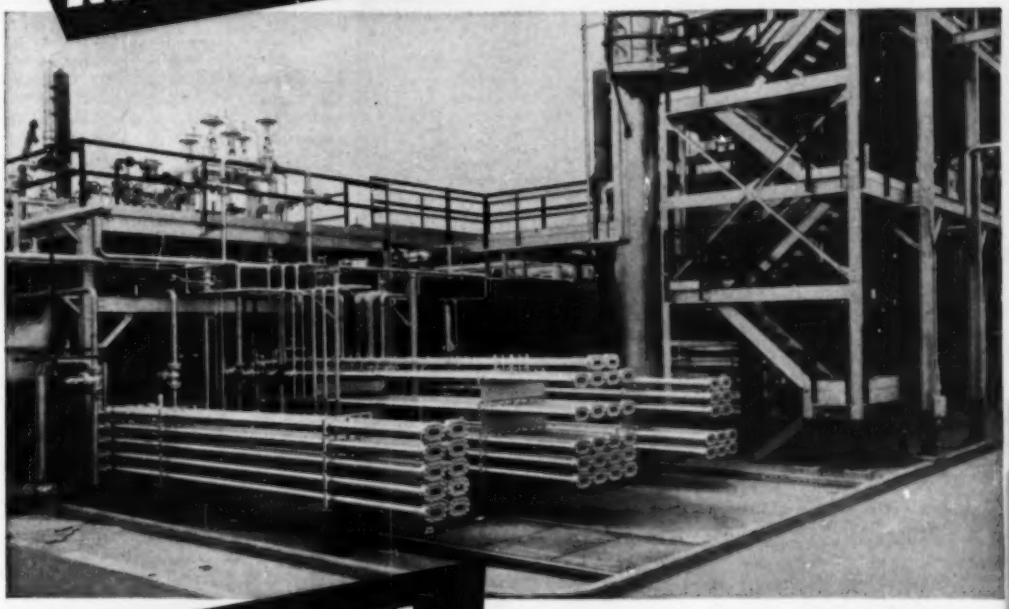

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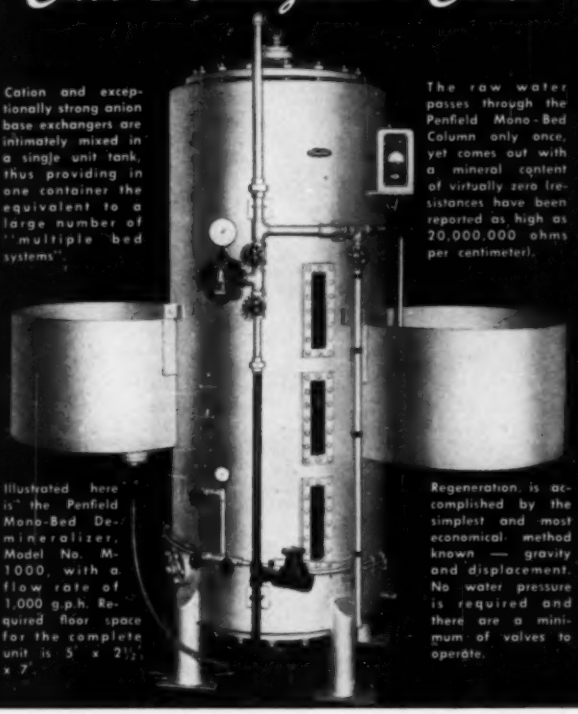
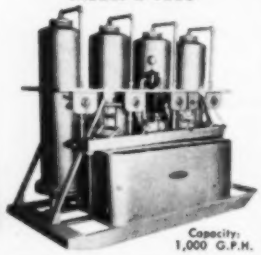
The raw water passes through the Penfield Mono-Bed Column only once, yet comes out with a mineral content of virtually zero (resistances have been reported as high as 20,000,000 ohms per centimeter).



Illustrated here is the Penfield Mono-Bed Demineralizer, Model No. M-1000, with a flow rate of 1,000 g.p.h. Required floor space for the complete unit is 5' x 2½' x 7'.

Regeneration is accomplished by the simplest and most economical method known — gravity and displacement. No water pressure is required and there are a minimum of valves to operate.

Model L-1000



Capacity: 1,000 G.P.H.

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Secretary's Report

S. L. TYLER

THE Executive Committee met at the Hotel Sheraton, Rochester, N. Y., Sept. 16. The first order of business was the approval of Minutes, receipt of the Treasurer's report and approval of bills. All candidates whose names appeared in the August Issue of Chemical Engineering Progress were elected to the grades of membership indicated with the exception of one who was held for further information. There were 11 Student members elected to membership: these applications had been received during the summer months.

H. T. Ward was appointed counselor of the student chapter at Kansas State College to replace W. T. Honstead. H. A. Neville was appointed counselor of the student chapter at Lehigh University to replace D. E. Mack.

Advices regarding a change in printing policy of The Institution of Chemical Engineers (London) were reported and complete details will be found on page 24.

Names of Emil Christofano, Paul J. Deutschman and E. Daniel Sadowski were added to the Suspense List which carries the names of those members who have entered the Armed Services.

The following additions were made to Committee Personnel:

Membership Committee — Kenneth R. Burke to succeed C. C. Long representing the Pittsburgh Section; George A. Schwabland representing the National Capital Section. In accordance with the established practice that where sections are being formed and chemical engineering clubs have been established preliminary to application for local section status, the following representation was approved: Connecticut—David Fields, Terre Haute, Ind.—Bernard L. Lubin, Michigan—Max Key.

The September Meeting of Council was held at the Hotel Sheraton Sept. 16. This was an important meeting of Council as advantage was taken of the attendance of committee chairmen at the Rochester meeting and interim reports were received from W. T. Nichols, chairman of the Committee on the Future of the Institute, G. G. Brown, chairman of the Chemical Engineering Education and Accrediting Committee, and H. G. Donnelly, chairman of the Professional Guidance Committee. These reports showed that the committees had made real progress during the year. Arrangments were made for further study by the Committee on the Future of the Institute and recommendations will be forthcoming in the near future. The Chemical Engineering Education and Accrediting Committee re-

(*Continued on page* 44)

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HASTELLOY alloy expansion joint

This HASTELLOY alloy C expansion joint is constantly exposed to acetic acid vapors saturated with acetyl chloride and containing small amounts of chloracetic acid. Alloy C is practically unaffected by these corrosive vapors at temperatures of 158 to 284 deg. F. Used at the top of a 24-foot distillation still, the joint is also elastic enough to compensate for the almost half-inch growth of the heated tower.

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SECRETARY'S REPORT

(*Continued from page 42*)

ported and made recommendations regarding action to be taken on ten chemical engineering curricula which had been inspected recently. The Professional Guidance Committee had several matters which subcommittees are studying and recommendations may be expected before the end of the year.

Letter had just been received from the Student Chapters Committee pointing out the reduction in the number of students to be registered and bringing up the question of the continuation of Student Chapter News. The Council voted that this activity should be continued for the school year 1951-52.

A six-months' financial statement of the Institute and Chemical Engineering Progress was presented and received.

Resignation from Council was received from W. G. Whitman, due to his recent appointment as director of Research and Development Board making excessive demands upon his time. B. F. Dodge was appointed to serve in his place for the balance of 1951.

The Secretary presented a report on members of the Institute whose membership had been terminated due to delinquency. This loss amounts to only about 1½ per cent.

The President reported the appointment of the following as Tellers to count and report on the Nominating and Election Ballots for Officers and Directors: W. D. Kohlins, chairman, H. L. Malakoff, C. C. Sutton and M. M. Hoover.

Secretary reported estimated costs for the publication and distribution of a Directory for the year 1952 to all members. Following the practice established in many other societies it was voted that no charge should be made for the Directory but that it would be distributed upon request only.

Preprints—Heat Transfer Symposium Papers

A Symposium on Heat Transfer has been set up by D. L. Katz and A. S. Foust which will be presented at the annual meeting in Atlantic City. This symposium will consist of 16 papers listed below and is scheduled for both morning and afternoon sessions on Wednesday, Dec. 5.

Temperature and Velocity Distribution in Wake of a Heated Cylinder—Part II—V. J. Berry, D. M. Mason and B. H. Sage, California Institute of Technology, Pasadena, Calif.

Heat Transfer in Air from a Single Tube in a Staggered Tube Bank—N. W. Snyder, University of California, Berkeley, Calif.

Natural Convection Transfer Processes, 1. Heat Transfer to Liquid Metals and Non-Metals at Horizontal Cylinders—Seymour C. Hyman, Charles F. Bonilla and

(*Continued on page 46*)

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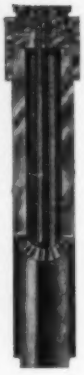

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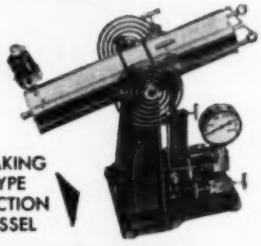

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On the basis of pilot plant performance, the company subsequently selected the Model 18-W-18-2 dual unit for a total capacity of 2000 GPH of water as the most suitable for their operation. In all, five such dual units were ordered. Each dual unit is installed on the outlet end of a demineralizer and has consistently produced water of fine clarity and brilliant polish, thereby increasing the reproductive quality of this company's TV screens.

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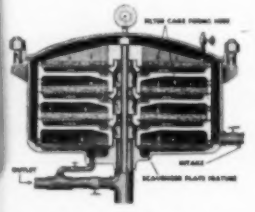
Manufacturers of fine filtration equipment for more than a quarter of a century.



Horizontal Plate Principle, unique Sparkler design, assures cake stability and uniformity. Cake cannot slip, crack or break even under intermittent flow.

Section of filter plate shows how horizontal design lends itself to high flow rates. Large, uncluttered channel between screen and supporting plate eliminates a principle cause of liquid flow frictions, permits free drainage.

SECRETARY'S REPORT

(*Continued from page 44*)

Stanley W. Enrlich, Columbia University, New York, N. Y.

Heat Transfer Coefficients for Liquid Mercury and Dilute Solutions of Sodium in Mercury in Forced Convection—T. C. Doody and A. H. Younger, Purdue University, Lafayette, Ind.

Performance of Small Liquid Metal Heat Exchangers—R. A. Tidball, Mine Safety Appliances Company, Callery, Pa.

Radial Heat Transfer Characteristics of Flowing Gas-Packed Bed Systems—J. R. Felix and W. K. Neill, University of Wisconsin, Madison, Wisconsin.

Heat Transfer Between Beds of Fluidized Solids and the Walls of the Container—Robert D. Toomey and H. F. Johnstone, University of Illinois, Urbana, Ill.

Thermal Design of Heat Exchangers Involving Three Fluids—V. Paschkis and M. P. Heisler, Columbia University, New York, N. Y.

The Effect of Forced Circulation Rate on Boiling Heat Transfer and Pressure Drop in a Short Vertical Tube—Joseph L. Schweppe and Alan S. Foust, University of Michigan, Ann Arbor, Mich.

Two-Phase One-Dimensional Flow Equations and Their Application to Flow in Evaporator Tubes—B. F. Harvey and Alan S. Foust, University of Michigan, Ann Arbor, Mich.

Boiling Coefficients Outside Horizontal Tubes—John E. Myers and Donald L. Katz, University of Michigan, Ann Arbor, Mich.

Solution of a Nonlinear Problem in Transient Heat Conduction Involving Temperature-Dependent Thermal Properties—John A. Beutler, Jr., and James G. Knudsen, Oregon State College, Corvalis, Ore.

External Heat Exchangers in Unsteady-State Systems—E. P. Lynch, Blaw-Knox Construction Company, Pittsburgh, Pa.

Heat Transfer to Slurries in Pipes, Chalk and Water in Turbulent Flow—Charles F. Bonilla, Armando Cervi, Jr., and Thomas J. Colven, Jr., Columbia University, New York, N. Y.

Heat Transfer Coefficients in Liquid Mixing Using Vertical Tube Baffles—I. R. Dunlap, Jr., and J. H. Rushton, Illinois Institute of Technology, Chicago, Illinois.

Thermal Conductivity and Prandtl Number of Air at High Temperature—Irvin Glassman, Princeton University, Princeton, New Jersey, and Charles F. Bonilla, Columbia University, New York, N. Y.

The above papers will be available about Nov. 1 at the Office of the Executive Secretary in a single volume on which the price is $2.50 postpaid. Please remit when placing order. Orders may be placed immediately by mail or on special order form which will be sent out with the annual meeting program. In addition preprints will be on sale at the Registration Desk at the Chalfonte-Haddon Hall in Atlantic City.

Because of time limitation it will not be possible to present these papers separately. They will, however, be presented in condensed form by the chairman of the session and then discussed. Therefore discussion will be based largely upon the preprints and those directly interested should obtain preprints.

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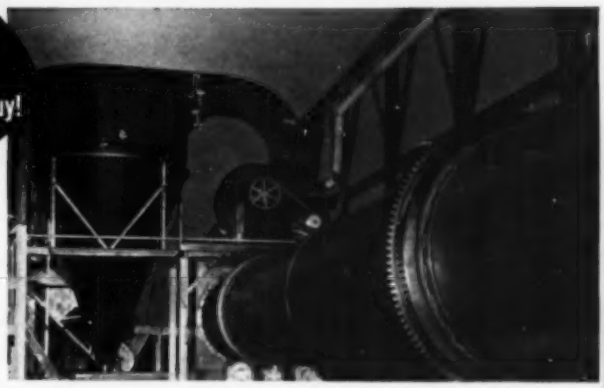

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News (continued from page 33)

INVITATION TO LEARN AT MUSCLE SHOALS

Chemistry and chemical engineering teachers from colleges and universities of the Midsouth will meet at Muscle Shoals, Ala., Nov. 9, in a one-day session to study production technology at the Tennessee Valley Authority's ammonia plant.

Sponsored by the departments of chemistry and chemical engineering of the University of Mississippi and the TVA's division of chemical engineering, this program is similar to the industry-sponsored school sponsored by the Chemical Education Projects Committee of A.I.Ch.E.

Frank A. Anderson, University of Mississippi, worked out the plan in cooperation with TVA for meetings where faculty members will observe plant operations and discuss process technology and problems with TVA engineers and chemists. The undertaking is part of TVA's program for making its research facilities and personnel available to colleges and universities of the region.

The program for the session will include such topics as history of ammonia synthesis, chemistry of the process, process variables and their effect, chemical control, techniques, design problems, process economics, and status of the industry.

All professionally interested persons are invited to be present. Additional information may be obtained from A. V. Slack, staff chemical engineer, Tennessee Valley Authority, Wilson Dam, Ala.

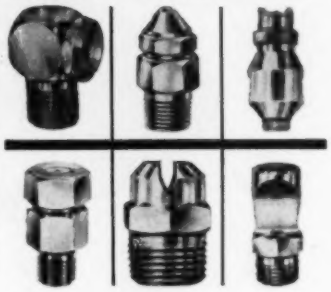
Editor's Note: This novel idea of the organization of schools, where faculty members can visit some industry, and chemical engineering teachers, for a day or two, can learn how the theories of chemical engineering are put to work, was explained at length in "C.E.P." September, 1951, p. 48, by Kenneth A. Kobe, Chairman, Chemical Engineering Education Projects Committee, A.I.Ch.E. This industry-sponsored school with its slogan, "teach the teachers," is catching on and according to certain prophets will be a boon alike to student and instructor.

NEW RESEARCH GRANT TO VIRGINIA TECH.

A new grant of $9,300 has been received by Virginia Polytechnic Institute, Blacksburg, Va., to continue a study for the air materiel command on plating of aluminum on steel. The first grant for this project, begun in April, 1950, was for $4,300. Nelson F. Murphy, an A.I.Ch.E. member and research professor of chemical engineering at Virginia Tech, is in charge of the fundamental research project first started nearly a year and a half ago.

(More News on page 50)

For chemical processing...



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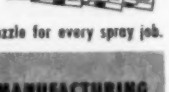

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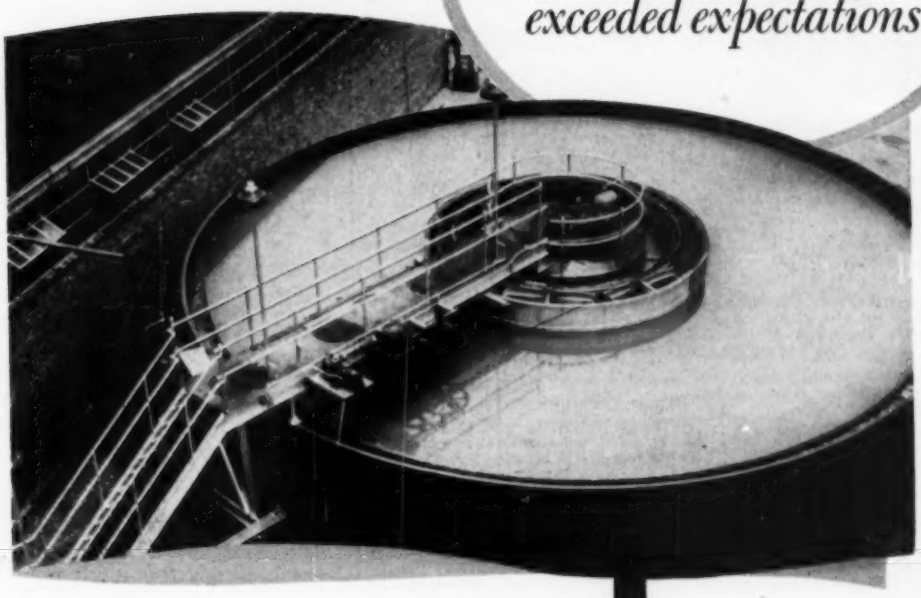

THE PROBLEM

Treatment of asbestos fiber wastes from mill board, insulation and wool felt products containing varying amounts of talc, carbon black and scrap material. Problem further complicated by shock loads occasioned by dumping of beaters without warning, which increased per cent of solids 2 to 3 times.

THE SOLUTION

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THE RESULTS

Suspended solids reduction ranging from 94.5% to 99.7%. Sludge concentration is easily held to between 1.5 and 2.5% by weight of solids which can be conveniently handled. Higher concentrations are possible, up to 6% having been obtained. Effluent is re-used for process water. Grit is removed from fibers, so that sludge can be re-used in mill.

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and painstaking research are responsible for the ability of Infilco equipment to handle really tough jobs, and contribute to cost-saving and space-saving superiority on any job. That's why outstanding concerns throughout the world invest in Infilco know-how and experience for the answer to their water conditioning, waste and sewage treatment problems.

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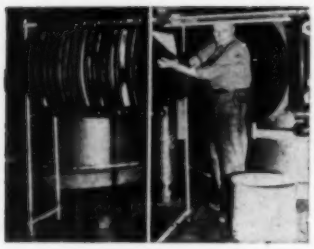
•

Produces Better Lard



with HERCULES No. 20 FILTER

A big meat packing concern installed a Hercules #20 Pressure-Leaf Filter (200 square foot capacity) to purify lard in a plant that produced 200,000 pounds of lard a day.

A saving of 8¢ per thousand pounds was obtained over the cost of filtering with the old type cloth and plate press method.

This means a saving was made of $16 per day, or $4,800 in a year of 300 working days.

The packer bought this filter because of the many advantages this closed filtering system gives in sanitation, cleanliness, and product quality. The money savings were a bonus.

- Filtration is finer, cheaper, and faster.
- Possible contamination is prevented by elimination of cloths.
- Cleaning reduced to 30 minutes of one man's time, once in 3 days —instead of 2 hours work by 2 men daily. One twenty-fourth the labor cost.
- Half the run is by gravity alone.
- No unfiltered lard is left in the filter at the end of a cycle.

•

Write for installation details—also for folders describing Hercules self-cleaning and manual Pressure Leaf Filters.

•

HERCULES FILTER CORP.

Manufacturers of filters for alcoholic and non-alcoholic beverages, sugar solutions, hydrogenated oils, fats, fish oils, gelatine, vinegar, cider flavors, pectin, etc.

204-21st Avenue Paterson, N. J.

NEWS

(*Continued from page 48*)

TELEVISION FOR MASS TRAINING

"Training by Television" is the title of a report prepared by the Navy Special Devices Center in cooperation with Fordham University. It presents the principal findings of an experimental study which compared training of Naval Reservists by live television with training by recorded TV and by standard classroom procedures. The verdict was television is a key to the rapid training of large groups of people. The report details the procedures followed to keep the tests objective. It concludes with a list of fundamentals that must be met if television is to be successfully exploited for rapid, mass training. This 24-page illustrated publication is available in mimeograph form at 75 cents each to the public through the Office of Technical Services, U. S. Department of Commerce, Washington, D. C.

CHEMICAL PROFESSIONS REPRESENTED AT DINNER

The Third Annual Dinner of the chemical professions in Cleveland will be held Nov. 8 at the Hotel Hollenden. Paul V. D. Manning, vice-president of International Minerals and Chemicals Co., and Director of A.I.Ch.E. (1949-51) will deliver the address.

This is an annual non-profit affair sponsored by the local chapters of the American Institute of Chemical Engineers (Cleveland Section), Electrochemical Society and American Institute of Chemists and American Chemical Society for the advancement of the chemical professions in the northern Ohio area.

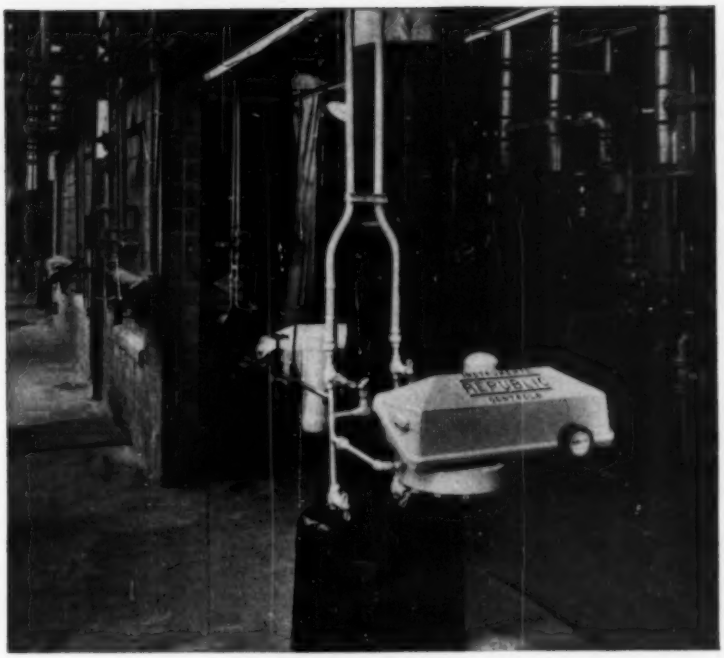
As vice-president in charge of research of International Minerals, Dr. Manning is responsible for all process development, equipment selection and design for all its plants producing many organic and inorganic chemicals used in fertilizers, food, pharmaceuticals, etc.

HARTLEY HONORED AT LUNCHEON

A luncheon in honor of Sir Harold Hartley, President of the Institution of Chemical Engineers, London, was given by A.I.Ch.E. in The Chemists' Club, New York, Sept. 13. Those present including Sir Harold were:

A. J. V. Underwood, The Institution of Chemical Engineers, London.
Thomas H. Chilton, President, American Institute of Chemical Engineers. DuPont Co., Wilmington, Del.
C. R. DeLong, Treasurer, American Institute of Chemical Engineers. North American Solvay Corp., New York, N. Y.
Donald B. Keyes, National Association of Manufacturers, Inc., New York, N. Y.
Paul D. V. Manning, International Mineral & Chemicals Corp., Chicago, Ill.
Irvin L. Murray, Union Carbide & Carbon Corp., South Charleston, W. Va.
R. Lindley Murray, Hooker Electrochemical Co., Niagara Falls, N. Y.
Earl P. Stevenson, Arthur D. Little, Inc., Cambridge, Mass.
Albert B. Newman, College of the City of New York, New York, N. Y.
Francis C. Frary, Aluminum Company of America, New Kensington, Pa.
John V. N. Dorr, The Dorr Co., New York, N. Y.
Joseph C. Elgin, Princeton University, Princeton, N. J.
John H. Perry, DuPont Co., Wilmington, Del.
Barnett F. Dodge, Yale University, New Haven, Conn.
Victor L. King, American Cyanamid Co., New York, N. Y.
Norman A. Shepard, American Cyanamid Co., New York, N. Y.
C. R. Downs, Consultant, New York, N. Y.
F. J. Van Antwerpen, Editor, Chemical Engineering Progress, New York, N. Y.
Stephen L. Tyler, Executive Secretary, American Institute of Chemical Engineers, New York, N. Y.

At luncheon in honor of Sir Harold Hartley—from left to right: Sir Harold Hartley, T. H. Chilton, A. J. V. Underwood, and C. R. DeLong.



These Meters Had To Be
ACCURATE—FLEXIBLE—RUGGED

One hundred twenty one Republic Pneumatic Transmitters are measuring the flow and level of various acids in a large chemical processing plant, a section of which is pictured above.

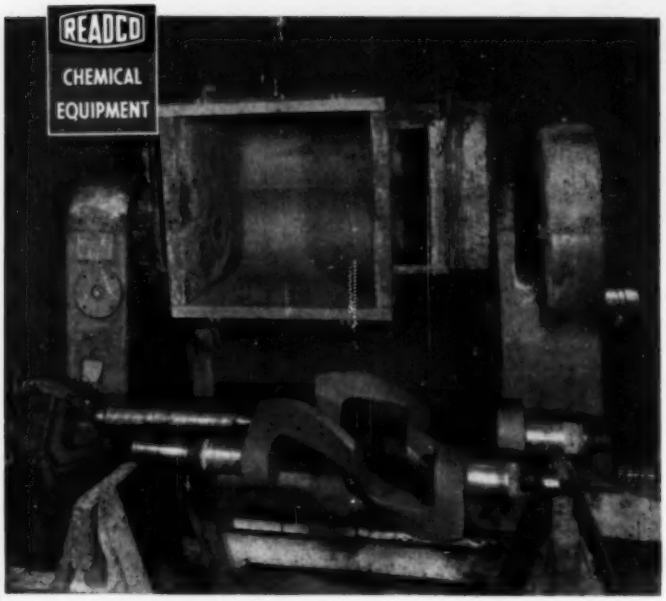
These transmitters have a guaranteed accuracy of within ½ of 1% of the transmitter range. By making a few minor adjustments or substituting a few small parts, their operation can be easily changed to any desired range between a minimum of 0—0.6" of water to a maximum of 0—704" of water. Their construction is more like that of a precision machine than of the sensitive instrument they are. It is for these reasons that Republic Pneumatic Transmitters, even though comparatively new, have already been specified and installed on over 2500 industrial metering applications.

Republic Pnuematic Transmitters are available for measuring flow, liquid level, pressure or liquid density of a wide range of fluids. They employ the force-balance principle to convert these process variables into air pressures which vary proportionally. These air pressures become direct measurements and can be conducted to reading instruments or used as the measuring impulse for the actuation of an automatic controller.

Data Book No. 1001 contains complete details on the construction and operation of Republic Transmitters. Write for your copy today.

REPUBLIC FLOW METERS CO.
2240 Diversey Parkway, Chicago 47, Illinois

After 10 YEARS of severe service mixing CORROSIVE BATTERY PASTE...



Each part of the Readco Double Arm Mixer shown above has been thoroughly tested and checked for indications of corrosion and wear after 10 years of severe service mixing battery paste. Upon completion of the examination, all drive parts have been found to be in excellent condition. The double arm, sigma blades need a minimum of remachining and replacement of original wear strips. A new bowl center and bowl ends are the only completely new parts needed.

Readco Double Arm Mixers are built to last, and they do it well. *Read Standard Corporation, Bakery • Chemical Division, York, Pennsylvania.*

FORMERLY Read Machinery Division of The Standard Stoker Company, Inc. ONLY THE NAME HAS CHANGED.

READ STANDARD
CORPORATION

ROCHESTER MEETING

(Continued from page 39)

W. T. Nichols, Monsanto Chemical Co., W. E. Gift, Tennessee Eastman Co.

The responses to his survey, he revealed, indicated that great improvements in technical education can be accomplished by increasing summer jobs for teachers, by increasing the number of graduate fellowships, and by providing more summer jobs for students.

In addition, it was indicated that greater use of teachers as consultants, providing one-year positions in industry for teachers, and more speakers for student gatherings, would aid in the education of students. Nichols said, ". . . as a start in effecting some of these improvements, the local section in St. Louis, sponsored practical lectures at Washington University, St. Louis, in chemical engineering.

Again Dr. G. G. Brown wound up the round-table conference by speaking on, "The Functions of Accrediting by the American Institute of Chemical Engineers." According to Dr. Brown, the A.I.Ch.E. was the first engineering body to accredit curricula of engineering colleges being inaugurated in 1923, first results being published in 1925. The number of schools meeting the standard at that time was 21% of the total. "The improvement since then," he said, "has resulted in having 60% of the technical colleges offering a chemical engineering curricula, accredited."

Maintenance Symposium

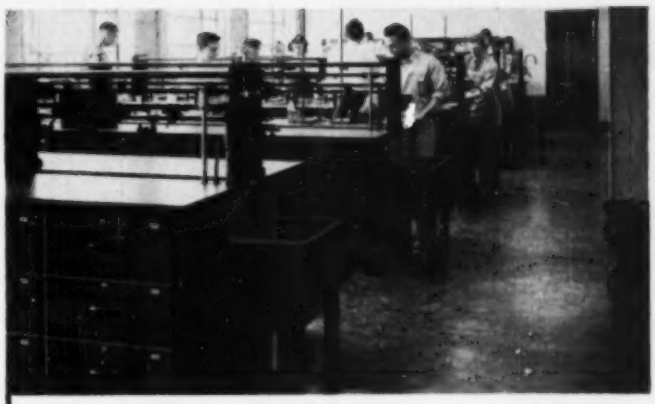
The maintenance symposium given the morning of the first day of the meeting stressed practical applications of maintenance theory and the functions of a maintenance organization in a chemical plant. According to L. A. Darling of the Du Pont Co. maintenance organizations in modern industry must be designed to fit specific needs and not be allowed to grow haphazardly. He talked on a new plan for organization and operation of maintenance in larger chemical plants, which offers a maximum of flexibility. He said, "Maintenance costs are rising and maintenance difficulties are beginning to appear as barriers to progress. Maintenance prob-

lems are becoming more important because of the increase in complexity and the high cost of chemical plant operation." He further stressed the importance of having maintenance areas of proper size under the supervision of an engineer with complete responsibility for maintenance operations.

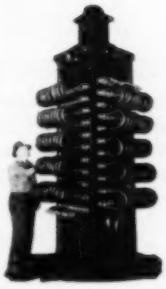
In speaking of the supply of engineers for maintenance jobs, he claimed that more talented engineers must be attracted to such careers. "If the problem is attacked effectively," he said, "the reluctance of young engineers to enter the maintenance field will be overcome." "The record," he continued, "seems to indicate that a young man starting in maintenance can hardly aspire to reach a high place in industry." He said it takes a good engineering education, character, energy, and interest in maintenance and years of experience in the field to develop a good maintenance supervisor, and some real inducement should be found to attract capable engineers to this field.

The part that intelligent maintenance plays in keeping the free enterprise system operating by a decrease in the cost of operation and keeping the margin between manufacturing and costs and sales prices, was developed by W. E. Chandler, superintendent of maintenance and construction of Monsanto Chemical Co. at Nitro, W. Va. His paper was on, "Preventive Maintenance in the Chemical Industry." Chandler felt that too little emphasis has been placed on the scientific approach to maintenance work. "Production plans of engineers to give efficient operation at reasonable costs," he said, "with the use of labor-saving devices and required equipment to increase yield, often requires the use of increased capital. Maintenance cost deduction has no theoretical limit and in many cases does not require capital. Preventive maintenance is an item which receives more and more recognition," he continued. "The fix-it-before-it-breaks-plan," he said, "has many economic advantages over the conventional fix-it-after-it-breaks-system.

Aside from specific suggestions about installation, operation, and maintenance of plant mixing equipment, the paper by R. D. Boutros, chief engineer of Mixing Equipment Co., Inc., in the maintenance symposium included many practical aspects of maintenance philosophy. According to Boutros, "easy installation, trouble-free operation, and low-cost maintenance in chemical plants, must be founded in good design," and he added that this constitutes the test of a good design. He further stated that a maintenance department is not alone responsible for keeping a plant running, and that even the best maintenance crew has trouble keeping a poorly designed or cheaply built unit on stream.



Corrosion Proof Equipment for Laboratory, Pilot Plant or Full Scale Production



For over 45 years the Knight organization has been producing equipment for use in corrosive service. Constructed of materials chosen to meet individual service conditions, Knight equipment includes:

Knight-Ware—acid and alkali proof chemical stoneware used in sinks, drain lines, fume ducts, containers, filters, valves, etc.

Permanite—a furan material reinforced with chemical glass fiber. Used as a construction material in HCl Absorbers, fume washers, heat exchangers, light-weight pipe and fume ducts.

Pyroflex—a thermoplastic lining material.

Pyroflex Constructions — complete engineered functional units, as process tanks, towers and complete HCl Plants. Each unit incorporates materials chosen to meet individual operating conditions. Materials may include glass, fused silica, plastics, carbon, lead, Knight-Ware, Pyroflex, Permanite and others.

Whether you need laboratory, pilot plant or production equipment, it will pay you to discuss your needs with M. A. Knight. When making inquiry, please specify equipment and service conditions. Literature will be sent on request.

Maurice A. Knight 710 Kelly Ave., Akron 6, Ohio
Acid and Alkali-proof Chemical Equipment

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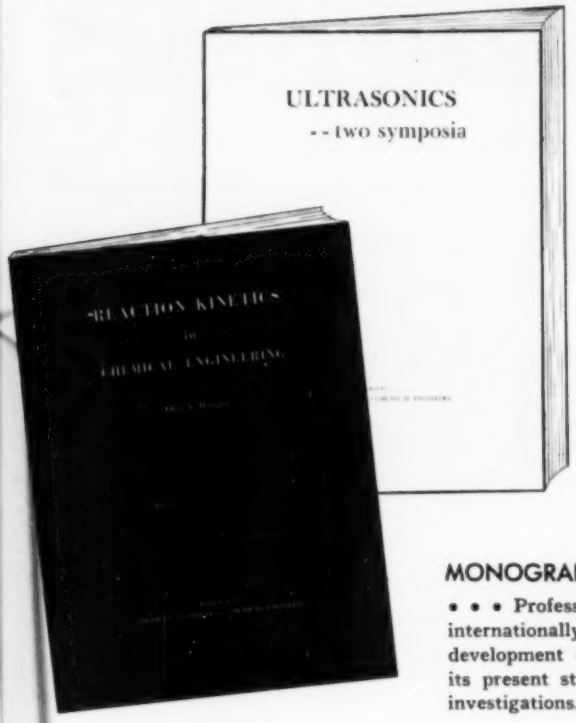

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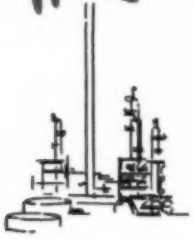

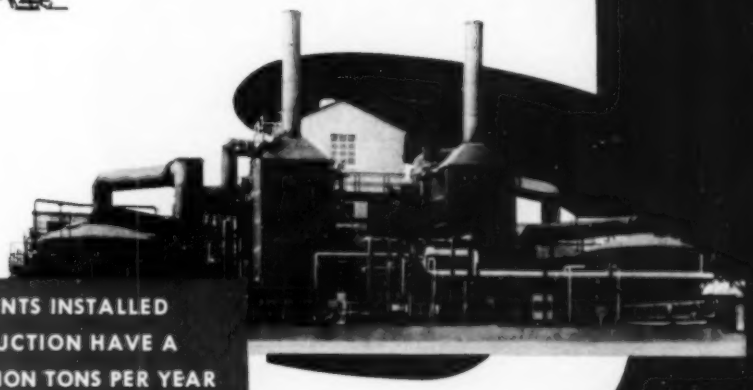

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This recent development, in which Foster Wheeler has functioned so prominently, promises to help bridge the present critical sulfur shortage.

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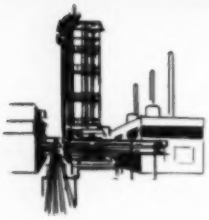

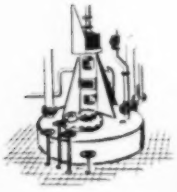

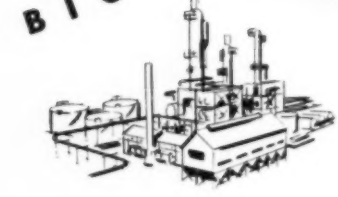

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DOWTHERM HEATING AND COOLING

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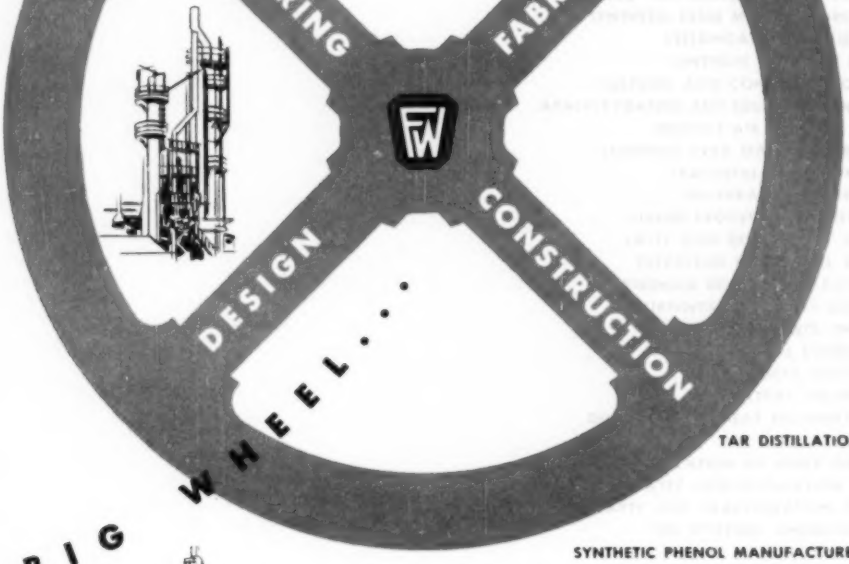
DEODORIZATION OF EDIBLE OILS

DIGESTER BLOW HEAT RECOVERY



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LOCAL SECTION

NEW YORK

New Techniques and Devices Symposium

How chemical engineers can apply effectively modern techniques and means for process engineering will be demonstrated at this section's annual symposium to be held at the Hotel New Yorker Oct. 25, 1951. J. F. Skelly, of the M. W. Kellogg Co., chairman of the symposium at this one-day meeting, and assisted by A. G. Parker, of Foster Wheeler, co-chairman, has arranged for the presentation of six papers which will cover in a general way the selection and design of equipment for industrial applications and the economic factors involved.

In the morning three speakers will be heard in the North Ballroom: David B. Pall, president of Micro Metallic Corp., will give a paper entitled, "Porous Metals," wherein he will discuss their manufacture from the raw-material stage and will include in his talk some technical considerations for process industry applications, such as commercial filters. D. Parker Litzenberg, president of the Atlantic Pump Corp., will talk on the Hermetik pump which was developed for special locations where conventional centrifugal pumps require frequent maintenance. "The Application of Automatic Electronic Computing Machines to Chemical Process Design Calculations," by William P. Heising and Dr. C. C. Hurd, associated with International Business Machines Corp., will cite the proved applications of punched-card computing machines in the chemical industry.

After luncheon the symposium will continue first with Robert Kunin, who will deliver a paper, "Recent Developments in the Industrial Applications of Ion Exchange." Dr. Kunin is associated with Rohm & Haas. After listing the various fields where large-scale operations of ion exchange are being conducted he will cite recent advances in ion exchange. J. M. Whalen, of the Trane Co., will discuss "Air-cooled Exchangers" and will acquaint process engineers with a new approach to the solution of cooling problems. "Flotation Chemicals," by J. A. Barr, Jr., of Armour & Co., will conclude the symposium with a coverage of flotation techniques in a unit operation of interest to the chemical engineer.

Registration for the meeting will commence at 8 A.M. in the Grand Ballroom lobby. Fees are as follows:

Symposium Session

A.I.Ch.E. Members	$2.00
Non-Members	3.00
Students	1.00

(*Continued on page 58*)

THESE ARE STRAIGHT FACTS ON SHRIVER FILTER PRESSES

- Lowest in cost per square foot of filtering area.
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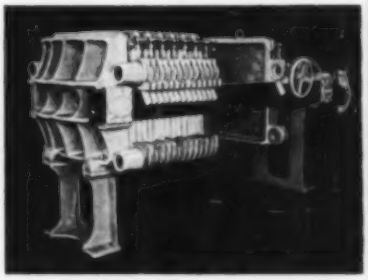
If the slings and arrows that are hurled at the head of the filter press were to find their mark, the poor old boy would have been a little-mourned corpse long ago. But strange to say, filter presses—and Shriver makes the major percentage of them—constitute the largest selling pressure filtration equipment in the process industries today and are likely to continue so for many years to come.

Despite claims for other filter types as to better filtering rates, faster operation, easier cleaning, etc., the facts prove the filter press a better overall investment, doing its consistently good job in more places on a greater variety of products under a broader range of operating conditions than any other filtering equipment ever developed. That's worth thinking about.

Filter press with special reinforced perforated plates and wire screen to assure free drainage, for filtering film dope at 250 p.s.i. pressure, equipped with Shriver "Hydro-Kloser."



Large filter press with stainless steel plates and frames, side feed, closed visible discharge, washing type.

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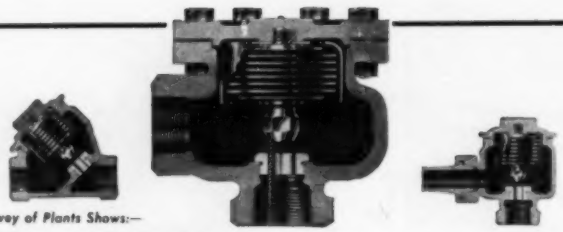

Survey of Plants Shows:—

15 of 20 Engineers Prefer Nicholson Traps

To determine the best steam trap on which to standardize, a large processing firm recently asked their plant engineers for their preference. In 15 out of 20 plants the choice was Nicholson.

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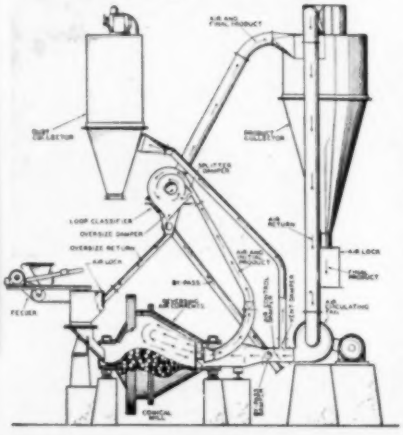

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NEW YORK

(*Continued from page 57*)

Luncheon

Registrants	$4.00
Non-Registrants	5.00

The first meeting of the 1951-52 season was held Sept. 25 at the Brass Rail Restaurant. Kenneth H. Klipstein, director of the chemical division of the N.P.A., talked on, "Washington and the Chemical Industry." Mr. Klipstein is on leave from the Calco chemical division of American Cyanamid Co., where he has been assistant general manager and head of the technical department.

DETROIT

A meeting was held at the Engineering Society of Detroit on Sept. 12.

A talk was presented on "Platforming for Aromatic Production," by C. V. Berger of Universal Oil Products Co. "Platforming" derived its name from the use of a platinum catalyst to make high octane gasoline from low-grade naphthas. The process is now being used to produce benzene and other products in short supply from petroleum. Approximately 75 members and guests were present.

Reported by D. L. Miller

NATIONAL CAPITAL

Government-Industry Relations Symposium

In an attempt to promote a better mutual understanding of the practical aspects of business dealings between government and industry, this Section will present a special symposium in Washington, Nov. 19. Entitled "Government-Industry Commercial Relations," the program will consist of three spokesmen who will explore in detail this complex, highly significant problem, particularly as it relates to the chemical industry. The moderator of the evening will be Bradley Dewey, president of the Dewey & Almy Chemical Co., and Rubber Director in World War II. The three speakers and their discussion topics will be:

The Law and the General Accounting Office's Operating Policies—Frank H. Weitzel, assistant to the Comptroller General of U. S.

The Problems of Governmental Agencies in Securing Contracts with Industry—A. S. Alexander, Under-Secretary of the Army

Private Industry's Point of View—George T. Felbeck, vice-president of Carbide and Carbon Chemical Corp.

The forum will include a panel discussion from the floor. This forum, the first of the formal meetings to be held by this Section during 1951-52, will be preceded by a dinner. Further details and dinner reservations may be obtained by writing to the section secretary, David P. Herron, Atlantic Research Corp., 812 North Fairfax Street, Alexandria, Va.

The first fall meeting was held Oct. 1 Paul McKee, executive secretary of the Maryland Water Pollution Control Commission, discussed progress in cleaning up the Potomac River. Mr. McKee had a number of colored slides to illustrate the talk.

Reported by David P. Herron

PITTSBURGH

This Section will hold a meeting Nov. 7, 1951, at the College Club in Pittsburgh, Pa. R. M. Griffiths of Price Waterhouse & Co., Pittsburgh, will address the meeting. His subject is titled, "What the Chemical Engineer Should Know About Corporate Finance."

Reported by Hugh L. Kellner

ATLANTIC CITY

(*Continued from page* 18)

ery ingenuity will do to create smart new models out of old hats. Mrs. Dorothy Nugent and her daughter, well-known radio and TV performers, tell and show all about how to do it in "Hats —Created, Not Dated."

Just about the most fascinating craft still practiced in this automatic age is glass-blowing, and Tuesday morning the ladies get to see it in action when Mr. and Mrs. R. M. Howell will give a lecture-demonstration of Bohemian glass-blowing. Then in the afternoon, the program lists a discussion of another ancient art now more popular than ever —that of floral arrangement. Mrs. Heatherly Dutton, who has been honored throughout America for her work (and recently in Japan, where they take only baseball more seriously), is the lecturer.

In case any of the ladies get homesick for housekeeping and cooking and all the other domestic joys, there's a busman's holiday scheduled for Wednesday morning, when there will be a guided tour of the hotel kitchens to see food preparation and cooking on a grand scale.

Cards and tables will be continuously at the disposal of the ladies in the Solarium in Haddon Hall.

Information Centers

This year's information desk will be located in the English Lounge of Haddon Hall, in the same area with the registration tables. Press headquarters will be on the same floor in Committee Room *'A.'* Don't seek any information on plant trips. Since Atlantic City is a long way from any suitable plant sites for chemical engineers, such visits will be eliminated from this meeting.

An information desk for the ladies will be found in the sun porch in Haddon Hall.

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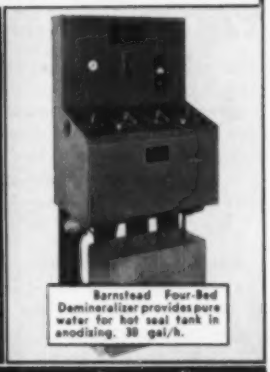
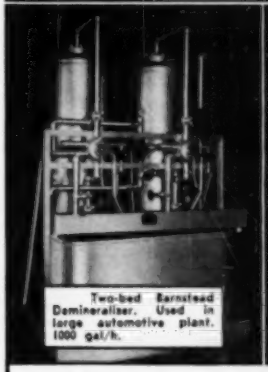


Barnstead Four-Bed Demineralizer provides pure water for hot seal tank in anodizing. 30 gal/h.



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FUTURE MEETINGS AND SYMPOSIA OF A.I.Ch.E.

Chairman of the A.I.Ch.E. Program Committee

Walter E. Lobo, The M. W. Kellogg Co., 225 Broadway, New York 7, N. Y.

MEETINGS

Annual — Atlantic City, N. J., Chalfonte-Haddon Hall Hotel, Dec. 2-5, 1951.

Technical Program Chairman: Frank J. Smith, Pan American Petroleum & Transport Co., 122 East 42nd St., New York 17, N. Y.

Atlanta, Ga., Atlanta Biltmore Hotel, Mar. 16-19, 1952.

Technical Program Chairman: H. E. O'Connell, Ethyl Corp., Box 341, Baton Rouge, La.

French Lick, Ind., French Lick Springs Hotel, May 11-14, 1952.

Technical Program Chairman: W. W. Kraft, The Lummus Co., 385 Madison Avenue, New York, N. Y.

Chicago, Ill., Palmer House, Sept. 4-6, 1952.

Technical Program Chairman: D. A. Dahlstrom, Chem. Eng. Dept., Northwestern University, Evanston, Ill.

Annual — Cleveland, Ohio, Hotel Cleveland and Carter Hotel, Dec. 7-10, 1952.

Technical Program Chairman: R. L. Savage, Dept. Chem. Eng., Case Inst. of Tech., Cleveland 6, Ohio.

Toronto, Canada. Royal-York, April 27-May 2, 1953.

Annual—St. Louis, Mo., Hotel Jefferson, Dec. 13-16, 1953.

SYMPOSIA

Opportunities in Sales for Chemical Engineers

Chairman: F. J. Curtis, Monsanto Chem. Co., 75 Rust Bldg., 1001 15th St. N.W., Washington 5, D.C.

Meeting—Atlantic City, N. J.

Chemical Engineering Fundamentals

Chairman: W. C. Edmister, Chem. Eng. Dept., Carnegie Inst. of Tech., Pittsburgh 13, Pa.

Meeting—Atlantic City, N. J.

Effective Speaking

Chairman: L. P. Scoville, Jefferson Chem. Co., 711 Fifth Ave., New York 22, N. Y.

Meeting—Atlantic City, N. J.

Vacuum Engineering

Chairman: W. W. Kraft, The Lummus Co., 385 Madison Avenue, New York, N. Y.

Meeting—French Lick, Ind.

Filtration

Chairman: F. M. Tiller, Dir., Div. of Eng., Lamar State College of Technology, Beaumont, Tex.

Authors wishing to present papers at a scheduled meeting of the A.I.Ch.E. should first query the Chairman of the A.I.Ch.E. Program Committee, Walter E. Lobo, with a carbon copy of the letter to the Technical Program Chairman of the meeting at which the author wishes to present the paper. Another carbon should go to the Editor, F. J. Van Antwerpen, 120 East 41st Street, New York 17, N. Y. If the paper is suitable for a symposium, a carbon of the letter should go to the Chairman of the Symposia, instead of the Chairman of the Technical Program, since symposia are not scheduled for any meeting until they are complete and approved by the national Program Committee. Before authors begin their manuscripts they should obtain from the meeting Chairman a copy of the Guide to Authors, and a copy of the Guide to Speakers. The first book covers the preparation of manuscripts, and the second covers the proper presentation of papers at A.I.Ch.E. meetings. Presentations of papers are judged at every meeting and an award is made to the speaker who delivers his paper in the best manner. Winners are announced in *Chemical Engineering Progress*, and a scroll is presented to the winning author at a meeting of his local section. Since five copies of the manuscript must be prepared, one should be sent to the Chairman of the symposium and one to the Technical Program Chairman of the meeting, or two to the Technical Program Chairman if no symposium is involved and the other three copies should be sent to the Editor's office. Manuscripts not received 70 days before a meeting cannot be considered.

PEOPLE

William C. Geer, who served as vice-president in charge of research and development at the B. F. Goodrich Co. until 1925, when he retired to private research, received the Charles Goodyear Medal for his contributions to rubber research at a dinner of the American Chemical Society's division of rubber chemistry held in the Hotel Commodore, New York, Sept. 6., as part of the Society's Diamond Jubilee meeting.

Recently Cornell University honored Dr. Geer by establishing the William C. Geer Laboratory of Rubber and Plastics. He was also honored in 1940 when he received a Modern Pioneer Award from the National Association of Manufacturers. Dr. Geer whose discoveries have netted more than 40 patents, received his Ph.D. in 1905.

Richard F. Shaffer, associate professor of chemical engineering at Pratt Institute, has been appointed head of the department of chemical engineering. Professor Shaffer, who has been on the Pratt faculty since 1945, is active in the New York Section of A.I.Ch.E., serving on the Symposium Committee and the New York Engineers Committee on Student Guidance. He is also a member of the American Society for Engineering Education. His industrial experience included work as research chemical engineer for the Du Pont Co., and as chief chemist for the Schlegel Lithographing Corp., for which he is now research director. He is consultant to a number of lithographing firms in the New York area, and teaches lithography in the division of general education at New York University.

Donald L. Katz, professor of chemical engineering of the University of Michigan, Ann Arbor, Mich., has been appointed chairman of the department of chemical and metallurgical engineering during the past summer. In June, 1933, he was employed by the Phillips Petroleum Co., Bartlesville, Okla. In 1936 he returned to the University of Michigan as assistant professor of chemical engineering. In 1942 he was made an associate professor and a year later professor of chemical engineering.

(*More about People on page 62*)

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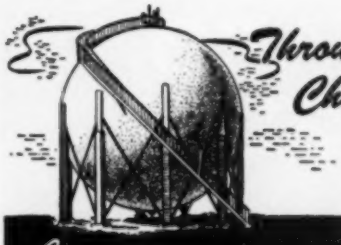

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Leo Garwin recently joined the research staff of the Kerr-McGee Oil Industries, Inc., in Oklahoma City. He was previously professor of chemical engineering at Oklahoma A. and M. College. He received his Ph.D. in chemical engineering from the University of Pennsylvania in 1947. At Publicker Industries, Philadelphia, from 1942-1945 his principal assignment was as shift supervisor in the butadiene pilot plant.

HOLLAND ON LEAVE WITH CHEM. DIV., N.P.A.

Christian V. Holland of Short Hills, N. J., was recently appointed deputy chief of the drugs, solvents and detergents branch, Chemical Division, of the National Production Authority, U. S. Department of Commerce.

Dr. Holland is on leave as coordinator of manufacturing for Sterling Drug, Inc., of New York. He started his career as a chemical engineer in the research and development laboratory of Arthur D. Little, Inc., of Cambridge, Mass., in 1925. He joined the Virginia-Caroline Chemical Corp., Richmond, Va., in 1928 and resigned in 1932 to become assistant director of the research and development laboratory of the Freeport Sulphur Co. in New York. Dr. Holland started with Merck and Co., Inc., of Rahway, N. J., in 1935, and resigned in 1947 as projects manager in the Foreign Relations Department to accept his present position with Sterling Drug, Inc.

Dr. Holland received his chemical engineering degree from the University of Virginia, Charlottesville, in 1921; the degree of master of science in chemical engineering practice in 1925 from Massachusetts Institute of Technology, Cambridge; and the degree of doctor of philosophy in 1924 from Johns Hopkins, in Baltimore.

Harold C. Weingartner has been elected vice-president and general manager of the equipment division of National Research Corp., Cambridge, Mass. Mr. Weingartner attended the University of Illinois where he received his B.S. in 1939. After receiving his M.S. in 1940, he was employed in the research department, Standard Oil Company (Indiana). From 1941 to 1945 he was associated with N.D.R.C. at the University of Illinois. He joined the engineering department of National Research Corp. in 1945, and became chief engineer in 1946. He held this position until his recent promotion.

WAGNER ADVANCED BY GENERAL ANILINE

Cary R. Wagner was recently elected senior vice-president of the General Aniline Works, the dyestuffs and chemicals division of the General Aniline and Film Corp., New York. He was previously vice-president in charge of operations.

Dr. Wagner joined General Aniline in 1947. His experience in the field of petroleum technology was turned to good account when he served as a member of the Petroleum Coordinator's staff during World War II. He is vice-president of the Synthetic Organic Chemical Manufacturers Association.

R. C. GUNNESS AIDS U. S. DEFENSE EFFORT

R. C. Gunness, a Director of A.I. Ch.E. (1951-53) and manager of research, Standard Oil Company (Ind.) has accepted for a period of one year effective Sept. 1 a position with the Research and Development Board, Department of Defense. His office is in the Pentagon, Washington 25, D. C. Mr. Gunness has been associated with the Standard Oil since 1938. He has been a group leader, section leader in the research department, prior to his assuming the duties of assistant director, associate director and then manager of research.

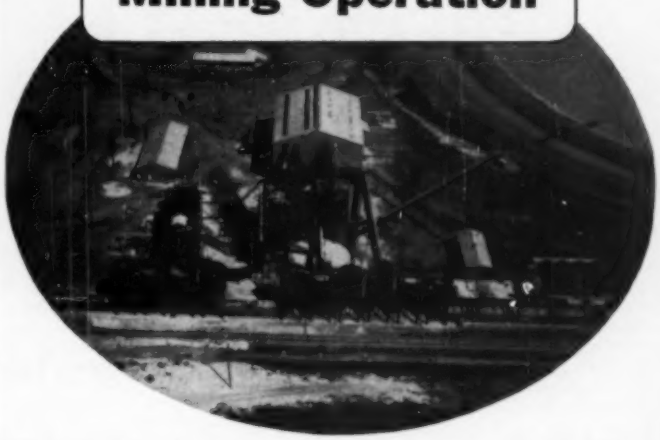
J. H. Zwemer of Sheffield, Ala., who has been project manager for Monsanto Chemical Co. during construction of the Chemical Corp.'s chlorine caustic plant at Muscle Shoals, Ala., is now plant manager for Monsanto which recently signed a letter contract with the Chemical Corps of the Department of the Army for the plant's operation and maintenance.

Nat Kessler was recently appointed to the newly created position of technical supervisor at the Painesville (Ohio) soybean processing plant of the A. E. Staley Manufacturing Co. Mr. Kessler, who has his bachelor's and master's degrees in chemical engineering from Washington University, St. Louis, Mo., has been a chemical engineer with the Staley company at Decatur, since 1944. He has been active in the Illinois Society of Professional Engineers.

Mr. Kessler had charge of the chemical engineering on a solvent extraction plant built by the Staley company in Decatur last year and participated in the preliminary engineering work in the Painesville modernization.

(*More About People on page* 65)

14 Horton Tanks Used in Florida Mining Operation



Late in 1949 the American Cyanamid Company placed its second phosphate rock mine in operation east of Tampa. This open-cut operation, known as the "Sidney" mine, helps supply hundreds of thousands of tons of phosphate rock products annually to agricultural and industrial users throughout the world.

Fourteen Horton welded steel tanks are used in a process that takes phosphate matrix from nearby deposits . . . breaks it up . . . washes, screens and deslimes it . . . and then recovers it for storage and shipment.

Four of the units we furnished are 50-ft. diam. processings tanks that deslime a fine slurry of sand, phosphate pebbles and mud, and hold it for transfer to the flotation plant. The other ten units are 22-ft. diam. hopper bottom tanks used to store both coarse phosphate pebbles and fine concentrate.

Specialized tank construction offers no obstacles to our design, fabrication and erection departments. Their knowledge enables them to tackle any problem with complete confidence.

Right: 10 Horton tanks used to store and load phosphate at the "Sidney" mine near Tampa, Florida. Four at the left are for coarse rock. Fine phosphate concentrates are stored in the other six.

The aerial view above also shows three of the four 50-ft. diam Horton tanks used to deslime slurry.

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Houston 2.....2154 National Standard Bldg.
Los Angeles 17...1541 General Petroleum Bldg.
New York 6.....3322—165 Broadway Bldg.
Philadelphia 3....1640—1700 Walnut St. Bldg.
Salt Lake City 4.....560 West 17th South St.
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PEOPLE

(*Continued from page 63*)

Alfred S. King has joined the staff of Wilson Meyer Co., West Coast distributors of Eastman industrial chemicals, manufactured by the Tennessee Eastman Co., a division of Eastman Kodak. Mr. King will supervise technical sales and service work on gasoline additives and other Eastman industrial chemicals. Formerly West Coast representative of the Houdry Process Corp., he has had extensive experience in petroleum and petrochemical processing.

William H. Muttera has been awarded an industrial research fellowship at Armour Research Foundation of Illinois Institute of Technology, Chicago, Ill. Muttera will do graduate work at the Institute and industrial research at the Foundation, each on a half-time basis. Completion of the work will lead to a master's degree in chemistry. Muttera is one of seven to win the fellowships, which were awarded on a competitive basis to holders of science degrees from accredited colleges. He received his bachelor's degree in chemical engineering in June, 1951, from Illinois Tech.

Frank Anthony Fazzalari, who received his M.S. degree from the University of Maryland in 1951 has joined the staff of the Central Research Laboratory of General Aniline & Film Corp. at Easton, Pa. He was formerly employed by the Civil Service Commission, Procurement Planning Section, Army Chemical Center, Md.

Rajendra Paul, teaching fellow at the Polytechnic Institute of Brooklyn, Brooklyn, N. Y., will continue his work for a doctor's degree in chemical engineering at the Polytechnic during this fall. During the past summer he worked in the pilot plant laboratories of the Foster Wheeler Corp. at Carteret, N. J. He is a native of India and has been in this country since September, 1948. He is co-author of "Distillation Equilibrium Data" published in 1950.

Jose B. Calva, technical director and president of J. B. Calva & Co., a consulting and research organization located in Minneapolis, Minn., addressed the Sept. 17 meeting of the Technical Association of the Fur Industry at the Governor Clinton Hotel in New York. Dr. Calva discussed the chemical and economic aspects of the industry, and stressed the outlook for a better fur industry through use of chemical engineering.

(*More about People on page 66*)

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CA Technical Topics

12-14% CHROME TYPE ALLOY

Norman S. Mott,
Chief Chemist and Metallurgist

This composition is a mildly corrosion resistant alloy capable of being heat treated for a variety of properties including high strength and hardness. It has a low coefficient of expansion, is tough, machinable, weldable, and finds extensive use in valve bodies, valve trim, pump parts, grinding units, etc., especially in power plants, oil refineries, and paper mills. The alloy has suitable corrosion resistance for applications involving superheated steam, hot oil, foodstuffs, and some neutral salts; however, it will rust to a slight extent in a moist atmosphere. Its low cost, high strengh, high hardness and wear resistance, coupled with an appreciable amount of corrosion resistance, make it an ideal alloy for specific applications where these factors are of primary importance.

This alloy is supplied in three carbon ranges: Under 0.15% C, 0.15 to 0.25% C, and 0.25 to 0.35% C, these being designated as CA 14S, CA 14I and CA 14H respectively. The carbon range selected depends upon the final degree of hardness desired as well as other properties. Increasing carbon up to 0.25% increases hardness appreciably in the quenched, and quenched and drawn below 900 F. condition. Above 0.25% the increase in hardness is slight and these higher carbon contents, unless the slightly greater hardness is very essential as in wear applications, are not generally recommended, as ductility and shock resistance are seriously lowered.

In the quench hardened condition this alloy has its best corrosion resistance. Drawing the quench hardened material at 600 F. increases the ductility and shock resisance without decreasing the hardness to any appreciable extent, and without any effect on the corrosion resistance. It also is very effective in eliminating hardening stresses and the danger of cracking, and so is highly recommended.

The best hardness range for machinabiliy is 220-240 BHN. This range is best obtained by tempering at 1200 F. for under 0.15% C, at 1300 F. for 0.15 to 0.25% C, and at 1400 F. for 0.25 to 0.35% C. Softening to too low a hardness causes gumming and poor machinability; also it causes low maximum hardness upon rehardening.

Molybdenum in amounts of 0.4 to 0.6% is sometimes added to improve machinability and high temperature creep resistance. Copper is often added to the extent of 1.0% to improve corrosion resistance and is occasionally combined with molybdenum additions. Nickel in amounts of over 0.75% and silicon over 1.0% are to be avoided. Nickel greatly increases the hardness in the tempering ranges used for machinability and necessitates lengthy special heat treatments. Silicon has a very detrimental effect on impact resistance when the alloy is drawn at 600 F.

Heat treatment for all carbon ranges of this alloy consists of oil quenching or air cooling from 1800 F. after slowly heating to temperature and holding for one hour, followed by tempering for 3 to 4 hours at the correct temperature necessary to obtain the desired properties. Tempering temperatures of from 900-1100 F. are to be avoided as this range produces low corrosion resistance and mechanical properties. Oil quenching from the hardening temperature produces slightly higher hardness values than air cooling, but is not practical for large castings. A hardening temperature of 1900 F. will produce slightly higher hardness values upon quenching in the 0.25 to 0.35% C grade.

Available on Request

Copies of this article in convenient filing form, and a table of mechanical properties for conditions of heat treatment most commonly used, are available. Address requests to Publicity Dept., The Cooper Alloy Foundry Co., Hillside, N. J.

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Be sure to stop off at the Chemical Exposition on your way to the Annual Meeting of the American Institute of Chemical Engineers at Atlantic City, December 2-5.

George F. Fitzgerald, associated with the Socony-Vacuum Oil Co., Inc., since 1925, will succeed Benjamin L. Harris who retired Sept. 1 as assistant superintendent of the Olean (N. Y.) refinery of the company after 41 years of service. Mr. Fitzgerald is a graduate of Notre Dame University. Prior to his present promotion he was supervisor of specialties at the local refinery of the company in Olean.

Mr. Harris started work in the company's Rochester laboratories following graduation from Amherst College in 1910. He became head of the transformer oil department in 1918 and went to Olean in 1923 where he assumed charge of the local refinery's transformer oil manufacture.

Clark W. Davis, assistant general manager of the Grasselli chemicals department of Du Pont, Wilmington, Del., since 1942, has been named general manager of that department. Dr. Davis, who has been with Du Pont since 1917, went abroad for the company in 1925 and 1929, traveling widely in the British Isles and Europe. Several patents have been issued in his name and in 1940 he received the Modern Pioneer Award.

John Vincent Bossi is now a chemical engineer for J. F. Pritchard & Co., Kansas City, Mo. Mr. Bossi, who received his B.S. in chemical engineering from the University of Kansas, was formerly with the Phillips Petroleum Co., Phillips, Tex.

Donald H. Buchanan, formerly a graduate student at Carnegie Institute of Technology, Pittsburgh, Pa., is now an engineer for the Esso Standard Oil Co., Baton Rouge, La. He received his B.S. in chemical engineering from the University of Pennsylvania.

C. C. Richiusa was recently named process manager in the engineering department of Monsanto Chemical Co.'s organic chemicals division at St. Louis, Mo. Mr. Richiusa, who received his engineering degree at Carnegie Institute of Technology, joined Monsanto in 1935 at Nitro, W. Va., and since 1942 has been in charge of all design work there.

William Budd is now a research engineer for the Goodyear Tire & Rubber Co., Akron, Ohio. He was formerly a graduate student of the University of Cincinnati, Ohio, and received his M.S. in chemical engineering from the University of Notre Dame.

HATCH, SELTZER WITH GENERAL ELECTRIC

Two chemical engineers have joined the Process Development Laboratory of the General Electric Co.'s chemical division.

Robert L. Hatch, a chemical engineer with the Standard Oil Company (Indiana) for three years, received a bachelor's degree in chemical engineering from Rensselaer Polytechnic Institute in 1945 and a master's degree from Massachusetts Institute of Technology in 1947.

Stanley Seltzer, a teaching fellow in the department of chemical and metallurgical engineering at the University of Michigan for two years, received his bachelor's degree in chemical engineering from Cooper Union in 1947, and his master's degree in 1948 and Ph.D. in 1951 from the University of Michigan.

The Process Development Laboratory is located at the division's headquarters at Pittsfield, Mass.

George M. Hebbard has been appointed deputy chief of the inorganic and agricultural chemicals branch of N.P.A.'s chemical division. Mr. Hebbard, who received his D.S. degree at the University of Michigan, is on leave as vice-president of operations and engineering of the Davison Chemical Corp., Baltimore, Md.

Louis E. Dequine, Jr., is now design engineer in the Nylon-Acrilan department of the Chemstrand Corp., Wilmington, Del. Mr. Dequine, who received his B.S. in chemical engineering from the University of Wisconsin, was formerly chief engineer of the American Bemberg Corp., Elizabethton, Tenn.

R. H. Morrison, Jr., was recently made production superintendent of the Holston Defense Corp., Holston Ordnance Works, Kingsport, Tenn. He had previously been assistant superintendent of Cellulose Esters Production, Tennessee Eastman Corp., Kingsport, Tenn.

Wayman L. Calhoun, formerly a junior process engineer for the Reynolds Metals Co.'s Hurricane Creek Plant at Bauxite, Ark., is now an instructor in chemical engineering at Rose Polytechnic Institute, Terre Haute, Ind. He received his M.S. from Massachusetts Institute of Technology, Cambridge, Mass.

William J. Baldridge is now with the Barrett division of the Allied Chemical and Dye Corp., Philadelphia, Pa., as a chemical engineer. He was formerly with the Borger refinery of the Phillips Oil Co.

Paul Freedley, formerly president of the Dunbar Corp., Dunbar, Pa., is now a staff engineer for the Diamond Alkali Co., Haddonfield, N. J.

(More about People on page 70)

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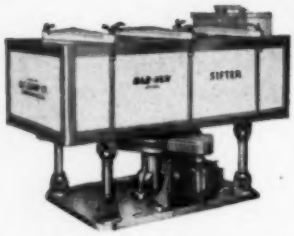

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CHEMICAL ENGINEER — B.S.Ch.E., 28, family. Experience in instrumentation, organic synthesis, and process development. Currently employed. Desire position with future in process development or related field. Box 5-10.

CHEMICAL ENGINEER — M.S., Columbia, 1949. Excellent scholastic record; Tau Beta Pi, Sigma Xi. Two years' experience in sales engineering and development. Single, draft exempt, age 25. Box 6-10.

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CHEMICAL ENGINEER—M.S., M.I.T., 1948. Married, 30, veteran. Five years in industry, including 3½ years supervision in chemical plants and in 8000 b/d integrated refinery. Desire production supervision in petrochemicals or related field. Box 15-10.

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CHEMICAL ENGINEER—B.S. Seven years background in petroleum refining, petrochemicals, ammonia, and synthetic fuels. Excellent experience in process design, economic evaluations, and correlation of pilot plant data. Desire position with aggressive oil or chemical company. Box 21-10.

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Nonmembers

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PEOPLE

(Continued from page 67)

Peter F. Mento, Jr., has joined the research and development staff of the Foote Mineral Co. as a research chemical engineer. He takes up his new assignment after a year with the Philadelphia branch, Quartermaster Corps. Prior to that he had been with the Du Pont Co. in its research and development division. Mento took his M.S. at the University of Delaware in 1949, and a B.S. in chemical engineering at Villanova in 1944.

Ray J. Stanclift, Jr., is now research engineer with the Carter Oil Co., Tulsa, Okla. Prior to this assignment he was graduate student and research assistant in the chemical engineering department of the University of Kansas, Lawrence, Kan.

John C. Harper is now on the staff of the University of California, Los Angeles, as associate professor, department of engineering. He was research engineer with California Research Corp.

Erwin C. Hoeman is now chief research chemist (chemical engineer) Marquette Cement Manufacturing Co., Des Moines, Iowa. He had been at Battelle Memorial Institute, Columbus, Ohio.

Walter F. Parker, Jr., is now chief chemist with the Dunn Paper Co., Port Huron, Mich. He formerly was in the capacity of process chemical engineer with Kimberly-Clark Corp., Meenah, Wis.

Necrology

ALBERT E. MARSHALL

Albert E. Marshall, a former President of A.I.Ch.E. and president of the Rumford Chemical Works from 1938 until 1948, died in Providence, R. I., Sept. 15. He had been consulting chemical engineer in that city since he resigned from the Rumford firm after its assets were sold to Hulman & Co., of Terre Haute, Ind. He was 68 years old.

Born in Liverpool, England, and educated at the Liverpool University, Mr. Marshall served as research chemist with the United Alkali Co., and as assistant manager of the Fleetwood Works, before coming to this country in 1912 after two years' service with the Thermal Syndicate, Newcastle-on-Tyne, to manage the latter's New York branch.

From then on he was engaged in many enterprises, such as his affiliation with the Davison Chemical Co. in Baltimore; as vice-president of Heyden Chemical Corp., New York; with the Aniline and Film Corp.; as chairman of the New England Interstate Water Pollution Control Commission, chairman of the Society of Chemical Industry, and as director of several projects. Another post he held was as a member of the advisory council of Princeton University department of chemical engineering and of the Engineering Experiment Station of Rhode Island College. As a consulting chemical engineer in Baltimore (1921 to 1928) and in New York (1928 to 1938) he specialized in the design of heavy chemical plant and process equipment, industrial uses of new construction materials, economic studies. In 1935 he assumed the position of vice-president of Amino Products Co. and in 1938 president of Rumford.

His A.I.Ch.E. activities included the following: Three terms a Director between 1916 and 1929; Chairman: Membership Committee (1929-31); Catalog Committee (1928); Patent Committee (1931-33) London Meeting (1930-36). He was also a member of the Publications Committee (1928-35); Institute's Representative American Engineering Council (1932); Vice-President, 1932, 1933 and President 1934, 1935, A.I.Ch.E. The Committee on Admissions (1938) and the Ethics Committee (1936-40, Chairman (1940) also engaged his attention.

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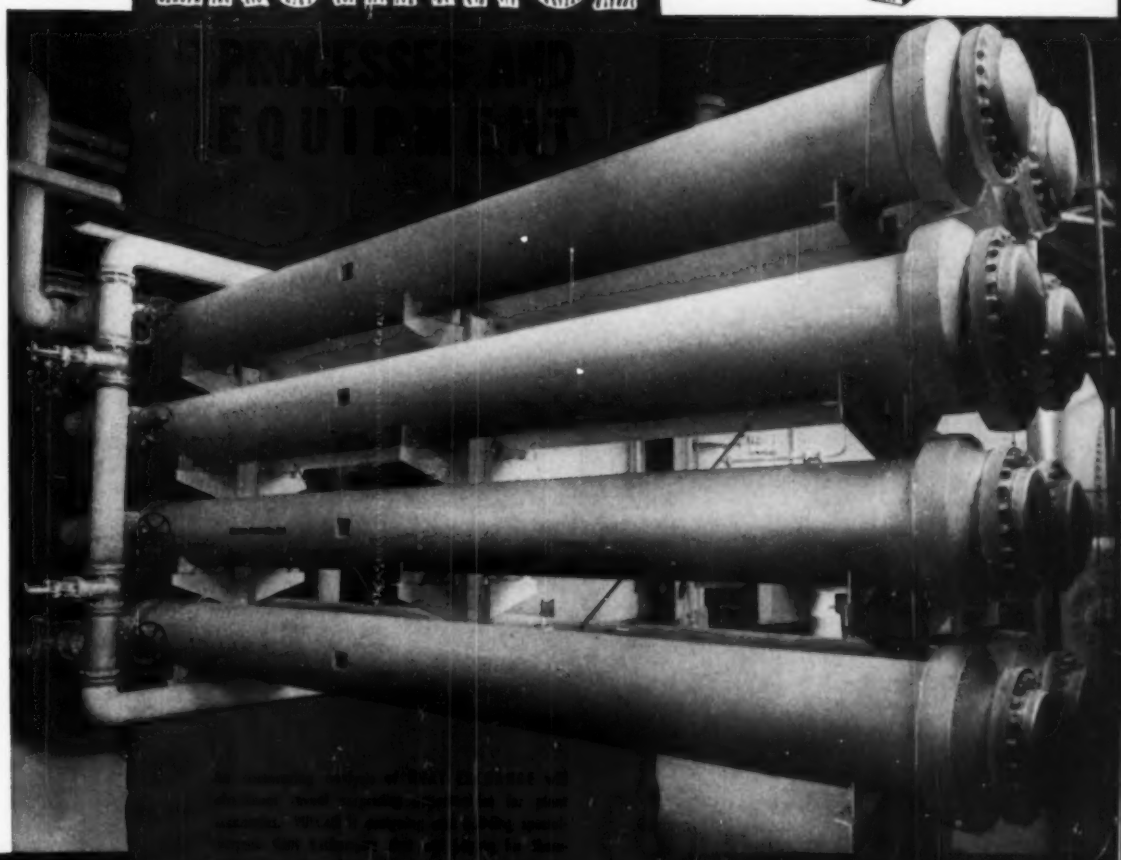

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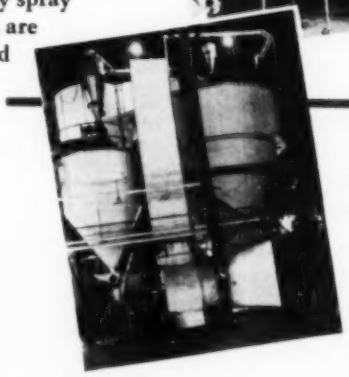

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